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ADVANCED IN-DUCT SORBENT INJECTION FOR SO_2 CONTROL

FINAL TECHNICAL REPORT

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ABSTRACT

The objective of this research project was to develop a second generation duct sorbent injection technology as a cost-effective compliance option for the 1990 Clean Air Act Amendments. Research and development work was focused on the Advanced Coolside process, which showed the potential for exceeding the original performance targets of 90% SO_2 removal and 60% sorbent utilization. Process development was conducted in a 1000 acfm pilot plant. The pilot plant testing showed that the Advanced Coolside process can achieve 90% SO_2 removal at sorbent utilizations up to 75%. The testing also showed that the process has the potential to achieve very high removal efficiency (90 to >99%). By conducting conceptual process design and economic evaluations periodically during the project, development work was focused on process design improvements which substantially lowered process capital and operating costs. A final process economic study projects capital costs less than one half of those for limestone forced oxidation wet FGD. Projected total SO_2 control cost is about 25% lower than wet FGD for a 260 MWe plant burning a 2.5% sulfur coal. A waste management study showed the acceptability of landfill disposal; it also identified a potential avenue for by-product utilization which should be further investigated. Based on the pilot plant performance and on the above economic projections, future work to scale up the Advanced Coolside process is recommended.

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BACKGROUND AND INTRODUCTION

In-duct dry sorbent injection technology has been actively developed in the U.S. since the early 1980s. The performance of these processes has been well-established through the development of the Coolside process (CONSOL)[1-6] and the HALT process (Dravo)[7,8] and through the DOE duct injection technology development program.[9-11] These development efforts included pilot-scale tests, proof-of-concept tests, and a full-scale utility demonstration. Established performance is in the range of 40-50% SO_2 removal at 2.0/1 Ca/S molar ratio and 20-25 °F approach to adiabatic saturation temperature using hydrated lime as the sorbent. Additionally, the 105 MWe demonstration of the Coolside process at the Ohio Edison Edgewater Station[4-6] showed that an SO_2 removal of 70% can be attained by improving the activity of the calcium hydroxide sorbent with sodium-based additive injection at a 0.2 Na/Ca molar ratio (~32% sorbent utilization).

Process performance data and economic analyses support the attractiveness of duct sorbent injection for a range of retrofit applications.[12] However, the applicability as a compliance option for the Clean Air Act or other regulations can be expanded by increasing SO_2 removals and sorbent utilizations. Higher SO_2 removals became more important for retrofit technologies with the passage of the 1990 Clean Air Act Amendments, which limit SO_2 emissions to 1.2 lb/MMBtu by the year 2000 and establish an emission cap thereafter. Higher sorbent utilization is an important area for improvement because of the large impact of sorbent cost on total SO_2 removal cost.

The objectives of the project entitled "Advanced In-Duct Sorbent Injection for SO_2 Control" (DOE Contract DE-AC22-91PC90360) are to improve the applicability of in-duct sorbent injection technology as a compliance option for the 1990 Clean Air Act Amendments and to reduce total SO_2 control costs. Specific desulfurization performance targets were established for realizing these objectives. These are to achieve 90% SO_2 removal and 60% sorbent utilization, while retaining the low capital cost and retrofit advantages inherent to in-duct sorbent injection technology. These targets represent a substantial improvement over existing sorbent injection technologies.

In Subtask 2.1 of this project, Evaluation of Advanced Concepts, pilot plant tests indicated that a process concept, referred to as the Advanced Coolside process, had the potential to achieve the process performance targets: 90% SO_2 removal and 60% sorbent utilization (Reference 13, Topical Report 1). Other concepts for advanced sorbent injection were evaluated in Subtask 2.1; however, none showed the potential to meet the process performance objectives. Therefore, the remainder of this project focused on developing and optimizing the Advanced Coolside process.

The Advanced Coolside process involves flue gas humidification to near the adiabatic saturation point using a contacting device which also removes fly ash from the flue gas. Downstream of the contactor, the sorbent (hydrated lime) is injected into the highly humid flue gas where it captures SO_2 before being removed in the existing particulate collector. The very high humidity promotes high SO_2 removal. High sorbent utilization is achieved by sorbent recycle. The recycle ratio can be higher than for existing duct sorbent injection processes because the fly ash is removed by the contactor prior to sorbent injection. Furthermore, addition of moisture to the recycle sorbent prior to reinjection significantly improves process performance.

Pilot plant development of the Advanced Coolside process was focused on the following areas:

- Optimization of sorbent recycle for improved sorbent utilization efficiency and increased SO_2 removal.

- Optimization of process equipment, for example, the contactor, for reduced capital and operating costs.

- Optimization of sorbent systems for improved performance and reduced cost.

- Evaluation of process operability issues.

- Evaluation of solid waste disposal and solid by-product utilization alternatives.

The results of process development in these areas are detailed in Topical Reports Nos. 2, 3, 4, and 5.[14-17]

Process conceptual design and economic evaluation was an ongoing activity during the development program. This allowed research and development to focus on approaches with the most potential for improving the process design and the process economics. A final report on the conceptual process design and economics is provided in Topical Report No. 6.[18]

CONCLUSIONS

1. The Advanced Coolside process achieved the process SO_2 removal target of 90% at sorbent utilization efficiencies over 70%, exceeding the original performance target of 60% sorbent utilization. The keys to achieving this performance were achieving near saturation in the contactor and optimizing sorbent recycle, including the moisture addition step. At a 1.2 fresh Ca/S ratio, a recycle ratio of 7 lb recycle/lb lime and a water addition level of 0.12 lb/lb recycle, SO_2 removal was 87% in the duct and 91% across the system.

2. Pilot plant tests showed that the Advanced Coolside process has the potential for very high SO_2 removal. Removals greater than 99% were achieved at sorbent utilization efficiencies exceeding 60%.

3. A second generation contactor and a third generation contactor were designed for significantly reduced capital and operating costs and smaller plant footprint compared to the initial design. Pilot plant testing confirmed performance in terms of humidification efficiency, particulate collection efficiency, operability, and desulfurization performance. The pilot plant testing also allowed optimization of operating conditions. The second generation contactor design consists of a short-residence-time spray chamber followed by a mist eliminator. The third generation design consists of a low-pressure-drop, in-duct venturi followed by a cyclonic separator.

4. Other process design and equipment improvements for reduced cost were realized through pilot plant development and engineering studies. These include improvements in the recycle wetting and handling systems, the sorbent preparation and handling system, the waste handling system, the flue gas reheat system, and the flue gas duct design.

5. Pilot plant tests of five different commercial hydrated limes, two specially prepared high surface area hydrated limes, and different quick-limes hydrated in a pilot hydrator showed that the Advanced Coolside process is relatively insensitive to the lime source. In conventional

sorbent injection processes, such as the Coolside process, sorbent source has a more substantial impact on performance.[4] The relative insensitivity of the Advanced Coolside process to sorbent source can be an economic advantage, allowing the use of the lowest cost sorbent available.

6. Pilot plant tests showed that the addition of small amounts of additives to the recycle sorbent during the water addition step can improve desulfurization performance in a baghouse. With the addition of NaCl or $CaCl_2$ at a ratio of 0.03 mol/mol fresh Ca, system SO_2 removals of 97% to greater than 99% were achieved at Ca/S ratios as low as 1.2.

7. Pilot plant operation provided a positive indication of the operability and retrofit potential of the Advanced Coolside process. Recycle test duration ranged up to 150 hr. A long-term (300 hr) performance test with 24 hr/day operation was conducted under optimized process conditions. No major operating problems were encountered. However, operability must be demonstrated in larger scale and longer term testing.

8. The waste management evaluation indicated that the combined spent sorbent/fly ash waste is suitable for landfill disposal. Further, the study indicated a potential for by-product utilization for synthetic aggregate production; a more thorough investigation of this potential is required.

9. A final process conceptual design and economic evaluation was conducted for the optimized Advanced Coolside process. The study projects capital costs less than one half of those for limestone forced oxidation wet FGD. The projected total SO_2 control cost on a levelized basis is about 25% lower than wet FGD for a 260 MWe plant burning a 2.5% sulfur coal. The levelized cost is sensitive to sorbent cost and, thus, is highly site-specific. This cost advantage meets previously established economic goals for increasing the attractiveness of the technology for electric utilities.

RECOMMENDATIONS FOR FUTURE DEVELOPMENT

Based on the process performance demonstrated in the pilot plant development program and on the favorable economic projections, scale up and further development of the Advanced Coolside process is warranted. Demonstration of the process on at least the 5-10 MWe scale is recommended to confirm the scale up of process performance and operability. Additionally, further process development is recommended in the areas of air toxics control and by-product utilization; development in these areas could give the process unique advantages over conventional FGD technology.

A number of technical issues need to be confirmed in scale-up tests before process commercialization. The key issues include:

- Scale up of pilot plant process desulfurization performance data and investigation of SO_2 removal potential of ESPs.

- Scale up of contactor performance and operability.

- Operability of the duct with sorbent injection into highly humid flue gas, including confirmation of the soot blower design included in the conceptual process design for preventing deposition.

- Operability of the existing ESP and the possible need for upgrade, in light of increased dust loadings and higher flue gas humidity.

- Operability of the recycle handling system, including moisture addition, transport and storage.

- Confirmation of performance during long-term continuous operation without shutdown periods.

- Confirmation of the sorbent injector design included in the conceptual process design for distributing the sorbent into the flue gas.

- Further investigation of recycle sorbent carbonation (and its effect on sorbent utilization) with an ESP.

- Further investigation on the effect of hold time on recycle sorbent activity.

The potential for by-product utilization should be explored further. The waste management study indicated that the potential exists for use of Advanced Coolside waste in producing lightweight synthetic aggregates for concrete masonry units. A more thorough evaluation of aggregate properties and a process design and economic evaluation are recommended. Utilization could have a significant economic benefit since waste disposal is a large component of the process cost. Since the aggregate market is a high-volume market, this potential use could give the Advanced Coolside process a unique advantage over existing technologies, for example, forced oxidation FGD producing gypsum.

The capability of the Advanced Coolside process to control air toxics should be investigated. If air toxics such as mercury are regulated, this capability would provide an added incentive to use the technology for SO_2 control. A literature analysis conducted by CONSOL suggests that the Advanced Coolside process has unique features for control of air toxics such as mercury and HCl. Both the gas/liquid contactor and the sorbent entrainment zone provide low temperature and efficient mass transport conditions important for capture of these species. Furthermore, the relatively high recycle ratios employed could increase the feasibility of using a more expensive co-sorbent such as activated carbon.

DESCRIPTION OF ADVANCED COOLSIDE PROCESS

Figure 1 shows a schematic of the Advanced Coolside process. The process achieves greater SO_2 removal and sorbent utilization than previous duct sorbent injection processes by operating at a higher flue gas humidity and by more fully exploiting the potential of sorbent recycle. The key to the process is a gas/liquid contacting device downstream of the air preheater. The contactor serves two purposes: to nearly saturate the flue gas with water and to remove most of the coal fly ash from the flue gas. The sorbent is injected downstream of the contactor into the highly humid flue gas. Hydrated lime is very active for SO_2 capture near the saturation point, even in the absence of liquid water droplets. Because the flue gas is already humidified prior to sorbent injection, there is no strict residence time requirement for droplet evaporation. SO_2 is removed by the sorbent in the duct and by that collected in the existing electrostatic precipitator (ESP) or baghouse. The heat of reaction between SO_2 and hydrated lime raises the temperature of the flue gas by roughly 8-10 °F for each 1000 ppm of SO_2 removed. Therefore, the particulate collector can be operated at an increased approach to saturation. However, because hydrated lime activity is highly sensitive to the approach to saturation, this reaction heat effect also acts as a limiting mechanism for SO_2 capture.

The spent sorbent is captured by the existing particulate collector as a dry powder. Sorbent recycle is an integral component of the Advanced Coolside process, allowing the sorbent utilization target of 60% to be exceeded. The potential for recycle is increased because fly ash is removed separately before sorbent injection. Furthermore, process performance can be improved by adding small amounts of water to the recycle sorbent prior to re-injection. The water acts to maintain a close approach to saturation by evaporating, thus, counteracting the heat of reaction. The moisture addition step is a key to maintaining sorbent activity and, thus, to achieving or exceeding the SO_2 removal target of 90%.

Equipment design optimization focused on the flue gas/water contactor. For the initial pilot plant tests the contacting device was a Waterloo scrubber.[13,14,19] In the process design optimization program, a second generation and a third generation contactor were designed, tested, and optimized. The improved contactor designs significantly reduce capital and operating costs. The third generation design consists of a low-pressure-drop, in-duct venturi followed by a cyclonic separator.

A more detailed conceptual process design for commercial application of the Advanced Coolside is provided in Topical Report No. 6 (Appendix A).[18]

EXPERIMENTAL

This process development program was carried out in a 1000 acfm (~0.3 MWe equivalent) pilot plant. Figure 2 is a schematic of the Advanced Coolside desulfurization pilot plant. It was designed to simulate integrated Advanced Coolside operation, including combined flue gas saturation and fly ash removal by a contactor, sorbent injection downstream of the contactor into the saturated flue gas, and steady-state continuous recycle with wetting of the recycle sorbent. The pilot plant, operating procedures, and analytical procedures are detailed in Topical Report No. 2 (Appendix B).[14]

In addition to pilot plant tests, some screening tests were conducted in a fixed-bed laboratory reactor, described in Topical Report No. 3.[15] Test sorbents were analyzed in a sorbent characterization laboratory described in Topical Report No. 3. The waste/by-product characterization study was conducted in a well-equipped waste characterization laboratory described in Topical Report No. 4.[16]

DISCUSSION

RECYCLE OPTIMIZATION

Optimization of sorbent recycle operating conditions was a key to exceeding process performance targets. Detailed data from this test program are presented in Topical Report No. 2 (Appendix B).[14]

Process Performance Goals Exceeded

The recycle optimization tests showed that the process performance targets of 90% SO_2 removal and 60% sorbent utilization could be exceeded. The 90% SO_2 removal target was achieved at sorbent utilizations over 70%. Very high SO_2 removal (90% to >99%) was achieved while maintaining at least 60% sorbent utilization. Sorbent recycle was a key to achieving these levels of performance. Figure 3 shows the SO_2 removals and corresponding sorbent utilizations achieved in the recycle optimization tests at different combinations of process variables.

The tests conducted with hot air reheat and with frequent baghouse pulse cleaning to simulate SO_2 removal with an ESP achieved 90% SO_2 removal at sorbent utilizations of up to about 75% (Table 1).

The tests conducted without reheat and less frequent baghouse pulsing to simulate SO_2 removal in a plant with a baghouse showed very high efficiency SO_2 removal (90 to >99%) while maintaining the target of 60% sorbent utilization (Table 2).

Effect Of Process Variables

Fresh Ca/S Ratio and Recycle Ratio. Increasing the fresh Ca/S ratio and/or the recycle ratio increases the amount of calcium available for reaction with SO_2. By maintaining a sufficiently high concentration of available calcium in the sorbent, the process target of 90% SO_2 removal can be achieved or exceeded. For example, at a 10 °F approach to saturation in the baghouse and with 0.15 lb H_2O/lb recycle, increasing the total available Ca/S ratio from 2.3 to 3.8 increased the in-duct SO_2 removal from 60% to 88% and the system SO_2 removal from 84% to 97%.

In-duct Residence Time. High SO_2 removals and sorbent utilizations were achieved with 1.7 to 2.0 s in-duct residence time. Below 1.7 s residence time, SO_2 removals were significantly lower. Above 2.0 s there was little effect of additional residence time on in-duct SO_2 removal.

Moisture Addition to Recycle. The addition of moisture to the recycle sorbent had a strong positive effect on desulfurization performance of the sorbent. For example, the addition of 0.15 lb H_2O/lb of recycle sorbent, at a 1.2 fresh Ca/S mol ratio, a 5.0 recycle ratio and a 10 °F approach in the baghouse, increased the in-duct SO_2 removal from 59% to 81% and the system removal from 73% to 88%. The sorbent utilization increased from 61% with no moisture addition to 71% with moisture addition. Moisture acts primarily to maintain a close approach to saturation by counteracting the reaction heat effect. Moisture also provides surface water to the sorbent particle which can enhance gas/solid reactions.

The optimum moisture addition level in the pilot plant tests was between 0.10 and 0.15 lb water/lb recycle sorbent. However, the optimum water addition level determined in pilot tests may not apply directly to large-scale operation because the ratio of transport air to sorbent is much higher in the pilot plant than a typical large-scale transport system, and the air used in the pilot plant is dry plant air. Consequently, in the pilot plant more water is required on the sorbent to allow for the evaporation into the dry transport air. The required level in a full-scale process will depend on the coal sulfur content, the extent of reaction, the design approach to saturation at the duct exit and the design recycle ratio.

Data Reliability

In the recycle optimization testing, there was good agreement between sorbent utilization based on the continuous flue gas analyzer and the fresh and recycle sorbent feed/composition data and the utilization based on baghouse solids analyses. There was no more than 4% absolute difference between the two values in any test. The agreement between the two values confirms the accuracy of process flow and analyzer data. There also was good agreement between the utilization calculated assuming steady-state recycle conditions and that based on baghouse solids analysis. The steady-state value is simply the system SO_2

removal divided by the fresh Ca/S mol ratio. There was no more than 5% absolute difference between the two values in any test. The absolute average of the differences was 2%. This agreement indicates that steady-state recycle conditions were closely approached. It further confirms the accuracy of the process flow and analyzer data. EPA Method 6 sampling tests were conducted during a recycle test; this confirmed the accuracy of in-duct desulfurization results.

EQUIPMENT DESIGN OPTIMIZATION

Equipment design optimization was a key to reducing the cost of the Advanced Coolside process. Descriptions of equipment designs and detailed test data are presented in Topical Report No. 2.[14] The focus of the design optimization was on the design of the contactor. Second and third generation contactors were designed that were mechanically simpler than the original design. Design optimization also focused on the sorbent recycle equipment.

Second Generation Contactor

Figure 4 is a schematic of the second generation contactor design, which consists of a short-contact-time spray chamber and a mist eliminator. It is substantially simpler than the initial contactor design (Waterloo Scrubber). One hundred fifty tests were performed to verify the saturation efficiency and to identify the optimum nozzle operating conditions for economic flue gas saturation and fly ash removal (Figure 5). Many operating conditions were tested that provided satisfactory saturation and ash removal. An optimum nozzle operating condition of 30 psig air pressure to each nozzle and 0.6 gpm/1000 acfm total water flow was chosen based on these results. The optimized operating conditions gave similar humidification performance and fly ash removals as the original design conditions for less operating cost; the lower cost is a result of the reduced air pressure (lower compressor capital cost and operating energy) and the reduced water flow (less pumping and waste water handling requirements). At ~730 acfm flue gas flow, nearly 100% relative humidity was achieved with 93% fly ash capture, using the optimized design conditions. The optimized contactor operating conditions were used in the subsequent recycle tests and in the sorbent optimization tests.

Contactor operability was good throughout the pilot testing. There were no problems with ash accumulation in the contactor, nor were there problems with mist eliminator plugging.

Third Generation Contactor

The third generation contactor (Figure 6) was designed for lower capital cost and a reduced plant footprint. It consists of a low-pressure-drop, in-duct venturi followed by cyclonic separator. Water is sprayed by hydraulic nozzles at the throat of the venturi, which reduces water droplet size and provides turbulent contact between droplets and flue gas for efficient particle capture and humidification. The water/fly ash mix is separated from the flue gas by the downstream separator. The design pressure drop for the venturi and separator and the design water requirement are higher than for the second generation contactor design; however, the third generation design is significantly smaller and has a significantly lower capital cost. Furthermore, the use of hydraulic nozzles instead of two-fluid nozzles save capital and operating costs for air compression. A detailed cost analysis is presented in Topical Report No. 6, Conceptual Commercial Design and Economic Evaluation.[18]

The pilot plant venturi contactor was purchased by CONSOL from Fisher-Klosterman. Initially, the venturi contactor did not achieve acceptable humidification efficiency, because of the very short contact time between the venturi throat and the cyclone in the small-scale unit. The contact time downstream of the throat is critical for humidification because the water droplet size is reduced in the throat. The contactor was modified to increase the contact time between the venturi throat and the cyclonic separator to about 0.1 s, which, in fact, better simulates the contact time of a full-scale unit. This increased residence time between the throat and the separator allowed reasonably close approaches (~1 to 4 °F) to be achieved. Injection of small amounts of steam at the exit of the cyclonic separator helped achieve near-saturation conditions (0 to 2 °F approach) for a wide range of flue gas flow rates. The steam injection rates ranged from 0.05 to 0.5 lb/min.

Fly ash collection efficiency of the venturi contactor was greater than 99% in four pilot plant tests conducted with EPA Method 17 sampling. This collection efficiency exceeds the target of 90% desired to reduce fly ash in the sorbent recycle loop. Fly ash collection efficiency was independent of gas flow over a range of 380 to 1025 acfm. The results indicate that a single venturi contactor can handle the range of turndown required for a commercial application to follow changing boiler load. This is an important result, since the first conceptual commercial design assumed that two parallel contactors would be required to handle load changes.

Operability of the third generation contactor was good throughout the performance tests. There were no problems with fly ash accumulation in the venturi, on the spray nozzles or in the cyclonic separator. There was a small amount of solids dropout in the horizontal duct upstream of the cyclonic separator; however, this accumulation leveled off after a short period of operation.

To meet performance targets, 5 to 15% more sorbent was needed with the venturi contactor than with the second generation contactor, presumably due to slightly lower humidity. The additional sorbent requirement was reduced to about 5% with the use of steam injection at the separator exit, with the use of supplemental nozzles or with slightly increased mist carryover from the cyclonic separator. This difference approaches the range of uncertainty in the pilot plant measurements. Based on the test results, the use of steam injection at the separator appears to be the preferred mode of operation.

Optimization Of Recycle Sorbent Treatment Equipment

A test was conducted in which the recycle sorbent was wetted using a pilot-scale, continuous pugmill. Performance of the pugmill was compared to the high intensity mixer used in previous pilot plant tests. The results from this test indicated that a pugmill can produce a satisfactory product, both from a materials handling standpoint and from a reactivity standpoint. These results are encouraging because a pugmill has substantially lower capital and operating costs than the high intensity mixer.

Other Design Optimization

In addition to pilot plant optimization testing discussed above, engineering studies were conducted to explore process improvements in all major process subsystems, including the sorbent handling, recycle handling, flue gas handling and waste handling systems. These engineering studies are discussed in detail in Topical Report No. 6, Conceptual Commercial Design and Economic Evaluation.[18] Key areas identified for process improvement/cost reduction include:

- Use of hydrocyclones instead of a thickener to concentrate the fly ash slurry before mixing with spent sorbent.

- Use of on-site lime hydration of quicklime for larger plants.

- Simplification of the flue gas reheat system.

- Improvements in the recycle handling system design.

- Simplification of the ductwork conceptual design.

SORBENT OPTIMIZATION

The Sorbent Optimization program included pilot plant evaluation of different sorbents including different commercial hydrated limes and specially prepared high surface area limes, an evaluation of hydration variable effects in a pilot hydrator, and testing of additive promotion. The results showed that process performance is relatively insensitive to hydrated lime source, compared to the conventional Coolside process. Small amounts of additives incorporated during the recycle wetting step were effective in promoting desulfurization, but only in the baghouse. Detailed test conditions and results of the Sorbent Optimization program are reported in Technical Report No. 3.[15]

Performance of Different Commercial Hydrated Limes

Pilot plant tests were conducted on five different commercial hydrated limes including Mississippi lime, the usual test lime in previous Advanced Coolside studies and in much of the development work for the conventional Coolside process. The limes tested were from different geographic areas and were selected from among the largest hydration plants in the country. The BET surface areas of the commercial hydrates tested ranged from 14 to 24 m^2/g. The desulfurization results showed only a small variation among the limes tested. From an economic standpoint, the relative insensitivity of the process to lime source can be advantageous, allowing the use of the lowest cost available lime.

Figure 7 shows once-through (no recycle) SO_2 removals at a 1.5 Ca/S mol ratio with the five commercial hydrated limes. In-duct removals ranged from 51 to 54%;

the system (duct + baghouse) removals were slightly higher with Mississippi lime, 80% versus 72-76% for the other limes.

Recycle tests were conducted with three of the commercial hydrated limes. These had a wide variation in surface area. At a 1.2-1.3 Ca/S mol ratio, a 6.8 recycle ratio (lb recycle/lb lime) and 0.12 lb water/lb recycle sorbent, the system (duct + baghouse) removals were very similar with all three limes (86 to 90%). In-duct SO_2 removals showed somewhat more variation, from 77% to 87%. Sorbent utilizations, based on analyses of baghouse solids, were near 70% for the three limes. Thus, a variety of commercial hydrated limes can be used to achieve the process performance goals of 90% SO_2 removal and 60% sorbent utilization.

<u>Performance of Alternative Sorbents</u>

Three specially prepared high surface area hydrated limes were tested in the pilot plant; surface areas ranged from 35 m^2/g to 41 m^2/g. Two of these samples were marginally more active than commercial hydrated lime in once-through tests; one sample was significantly less active than commercial lime, despite its high surface area. In recycle tests, the performance of the most active of these limes was not significantly better than commercial hydrated limes. Thus, their use in the Advanced Coolside process does not offer a significant advantage over commercial hydrated limes.

Once-through tests were conducted with a finely pulverized limestone. Although limestone is not a sufficiently active sorbent for commercial use in the Advanced Coolside process, the results indicated that $CaCO_3$ does have significant desulfurization activity. This may be an important observation, because some $Ca(OH)_2$ is converted to $CaCO_3$ in the Advanced Coolside process. The activity difference may be largely a result of the lower surface area of limestone (1.6 m^2/g) compared to that of hydrated lime.

<u>Pilot Plant Hydration Test Program</u>

An experimental program was conducted to determine the effects of selected hydration variables on hydrated lime properties and desulfurization performance. The testing was conducted in the Dravo Lime Company pilot hydration test facility. The hydration process variables studied were: feed lime particle size, water temperature, lime feed rate and target final product moisture content. The target product moisture is the projected moisture content of the product hydrate calculated from the stoichiometric water feed and the water temperature. Two different quicklimes were tested. After hydration, the products were analyzed for porosity, BET surface area, particle size and moisture content. Reactivity of the hydrates to sulfur dioxide was determined by two different techniques: a laboratory-scale utilization test and Advanced Coolside pilot plant once-through testing.

Detailed data from the hydration tests are presented in Topical Report No. 3.[15] Statistical analysis of these data and computer modeling studies using this data base are discussed in Topical Report No. 5.[17] Data analysis indicates that there is no significant correlation of desulfurization activity with sorbent physical properties or with quicklime source. Correlation equations were developed for desulfurization performance as a function of hydration variables. However, the tests indicated that the Dravo pilot plant hydrator did not closely simulate commercial hydration. Pilot produced hydrates had generally lower surface areas and lower desulfurization activity than commercial hydrated limes. Therefore, the correlations cannot be used for predictive purposes.

Additive Addition to Recycle

Recycle tests were conducted with small amounts of additives incorporated in the combined recycle and fresh sorbent during the moisture addition step. A moderate enhancing effect of small amounts (~0.03 mol/mol fresh Ca) of inorganic chloride compounds ($NaCl$, $CaCl_2$) on sorbent performance was observed in the baghouse but not in the duct. Therefore, use of small amounts of these additives may be an attractive means of achieving high SO_2 removal efficiencies (90 to >99%) in a plant with a baghouse. Additive incorporation in the recycle pretreatment step is attractive because it uses existing equipment and commercially available hydrated lime.

PROCESS PERFORMANCE TESTING

The objective of the performance testing was to generate performance and operability data for design and scale-up of the process. The performance test consisted of about one week of operation with two shifts per day followed by three separate weeks of 24 hr/day operation. The total on-stream time was 295 hr. The purpose of the initial week of testing was to establish near steady-state operating conditions and sorbent composition. The purpose of the around-the-clock operation was to evaluate performance and operability issues during longer periods of continuous operation. Although the test was divided into three periods of 24 hr/day operation, the same sorbent material was used; that is, the baghouse material collected at the end of one period was used as the recycle material at the beginning of the subsequent test period.

The test conditions for the performance test were selected based on the results of the previous process optimization tests.[2] The Ca/S ratio was in the range of 1.2 to 1.3 for the test. The recycle ratio was 7 lb/lb fresh lime and the recycle water addition level was about 0.12 lb/lb recycle. The third generation contactor (venturi + centrifugal separator) was employed for all the testing; it was operated to achieve near saturation (0 to 2 °F approach) conditions. For most of the testing the flue gas was reheated to give a baghouse approach of ~20 °F and the baghouse was pulse-cleaned continuously; this approximately simulated the SO_2 removal expected with an ESP. For part of the last test period, no reheat was employed; this maximized SO_2 removal in the baghouse.

Although the scale of the 1000 acfm pilot plant is not sufficient to fully resolve process operability issues, the performance test provided a positive indication of the operability of the Advanced Coolside process during relatively long periods of continuous operation. The key operability issues evaluated were operation of the flue gas duct with sorbent injection at high humidity and operation of the recycle sorbent wetting, handling and transport systems. These and other issues should be further evaluated in larger scale, longer term tests.

There were no major operating problems in the flue gas duct with injection of wetted recycle sorbent at high humidity. The duct was operated at an inlet approach of 0 to 2 °F and an exit approach of 5 to 8 °F. As discussed in a previous report,[14] the pilot plant had a duct configuration with numerous changes in flue gas direction, presenting more potential for operating problems than typical commercial systems. Because soot blowers are included in the conceptual process design developed in Task 5 of this project, the duct was periodically air lanced with 50 to 80 psig air to simulate soot blowing. The soot blowing was effective in preventing accumulation of solids in the pilot plant duct. The material which adhered to the duct walls was generally soft and easily removed and carried to the baghouse by the soot blowing. The soot blowing was used primarily at elbows and near the sorbent injection point. As observed previously,[14] the amount of accumulation in straight duct runs was small and tended to level off with time even without soot blowing.

For most of the performance test, the recycle sorbent was effectively wetted, fed, and transported to the flue gas stream. Recycle handling did, however, require frequent operator attention, although much of this attention was specific to the small scale and the specific equipment employed in the pilot plant. There were rather frequent instances of eductor plugging, a fairly common problem in small-scale systems, because the orifice in the eductor venturi is quite small. The problem was effectively managed by periodically cleaning the orifice of the eductor with a rod to remove deposits. It is anticipated that this would not be a significant problem with properly designed commercial-scale pneumatic transport equipment.

A system SO_2 removal of approximately 90% at 1.2 to 1.3 Ca/S was maintained during the performance test. In-duct SO_2 removal was lower (average, 78%) than previously observed at similar conditions (> 85%). This was partly a result of a higher approach at the duct exit (5 to 8 °F) than in previous tests (3 to 4 °F). The approach can normally be controlled by adjusting the amount of recycle water addition; however, the pilot plant was operated at the maximum operable water addition rate. The sorbent agglomeration caused by operating at the maximum amount of moisture addition may have also contributed to the lower duct removal.

A higher degree of sorbent carbonation was observed in the performance testing than in most previous testing. The $CaCO_3$ content of recycle sorbent ranged from 20 to 30 wt%. The degree of carbonation may be high in the pilot plant because the sorbent in the baghouse is intimately contacted by flue gas with a low SO_2 content and high humidity. The extent of carbonation and its effect on performance should be further evaluated in larger scale testing with an ESP.

The performance testing is discussed in detail in Topical Report No. 4.[16] Detailed operability and performance data are presented.

WASTE MANAGEMENT EVALUATION

The initial objective of the waste characterization study was to develop the data needed for designing the waste handling and disposal systems for the process. The waste characterization test program was expanded to include exploratory tests of by-product utilization options. This involved pelletization tests and preliminary evaluation for production of synthetic aggregate materials.

The Advanced Coolside process generates two waste streams: the dry spent sorbent from the particulate collector and the fly ash/water slurry collected in the contactor and subsequently concentrated. The proposed concept for disposal or utilization is to mix the two streams, controlling the overall moisture content by controlling the water content of the fly ash slurry.

Three Advanced Coolside waste samples were prepared for use in the waste characterization study. These samples represent simulated Advanced Coolside waste produced from a boiler using feed coals with 7.5% ash and 3.5%, 2.5% and 1.5% sulfur. Advanced Coolside waste samples were characterized to ensure that adequate information is available on the physical and chemical nature of the waste for the design and construction of safe and stable landfills. The properties of the waste characterized include composition, moisture and density relationship, unconfined compressive strength and leaching characteristics.

The maximum dry bulk density of Advanced Coolside waste increased from 75 to 80 lb/ft^3 with increasing fly ash component in the waste. The fly ash component in the waste increased with decreasing sulfur content of the coal from which the waste was generated. The moisture content which gave the maximum density (optimum moisture) was about 32% (dry basis).

Advanced Coolside waste, compacted to 95% of Proctor density and optimum moisture, has unconfined compressive strength that is suitable for landfill disposal. The strength increased from 20 psi (uncured) to 100 psi or more after 28 days of curing. As a point of reference for unconfined compressive strength values, a person walking exerts pressure of about 5 psi and bulldozers used in landfills exert pressures ranging from about 12 psi to about 19 psi.

The leachate toxicity of Advanced Coolside waste was determined. The leachates were prepared according to both the TCLP and ASTM leaching procedures. The trace element (As, Ba, Cd, Cr, Pb, Hg, Se and Ag) concentrations were well below (by at least a factor of 50) RCRA allowable limits. Thus, the waste can be classified as non-hazardous for landfill disposal. In addition, the concentrations of Fe, Mn, Ca, Na, Al, sulfate, K and total dissolved solids (TDS) in the leachates were similar to those from other dry flue gas desulfurization (FGD) wastes.

Pelletization takes advantage of the cementitious properties of the Advanced Coolside waste to make products that may be applicable for use as synthetic aggregates. Pelletization also can improve waste handleability and reduces waste leachability. The Advanced Coolside wastes were pelletized on a pilot-scale disc pelletizer. The pellets produced were lightweight and had low bulk specific gravity. The pellets also had a desirably low LA abrasion index, low water absorption, and a coarse size distribution; however, they also had a high soundness index (i.e, low durability). These data indicate that pellets made from Advanced Coolside wastes may have potential for use as lightweight coarse aggregates in concrete masonry units. For this use, there is no soundness index specification. A more thorough evaluation of other pellet characteristics for this application is recommended. An evaluation of potential economic impacts also is recommended.

The waste characterization study is discussed in detail in Topical Report No. 4.[16] Detailed data on waste properties are presented. Detailed data on the pellets produced from Advanced Coolside waste also are presented.

CONCEPTUAL PROCESS DESIGN AND ECONOMIC EVALUATION

The objectives of Task 6, Conceptual Design and Economic Evaluation, were to develop a conceptual design for a utility-scale application of the Advanced Coolside Process and to assess the economic attractiveness of the process. An additional CONSOL objective was to identify process areas for potential cost reductions to guide research efforts in areas that would most impact the economics and the commercial readiness of the process. As a result, engineering and economic evaluation commenced early in the project and was an ongoing process.

In early 1993, an interim process economic evaluation was completed. Results indicated that Advanced Coolside had an economic advantage relative to limestone wet scrubbing for a range of plant sizes and coal sulfur levels. The evaluation also identified several areas for potential process improvement, including equipment design optimization and sorbent utilization optimization. Areas identified for design optimization included improvement of the gas/liquid contactor design, improvement of the sorbent recycle handling system, and improvement of the waste handling system. Based on the results of the interim economic study, economic targets were established for the process. These were to achieve a 20% levelized cost advantage and a 50% capital cost advantage over limestone wet scrubbing for a range of plant sizes and coal sulfur levels. Based on conversations with utilities, these levels of cost advantage would make it attractive to consider a less-developed technology.

Topical Report No. 6 (Appendix A)[18] presents the results of a final conceptual process design and economic study for the Advanced Coolside Process, under DOE Contract DE-AC22-91PC90360. It describes a complete conceptual process design for full-scale, coal-fired applications of the process. Advanced Coolside process costs were compared to those of limestone forced oxidation (LSFO) wet FGD technology. The process economics were investigated for coal sulfur levels ranging from 1.0% to 3.5% (as-received) and plant sizes ranging from 160 to 512 gross MW. The final economic study incorporates the results of pilot plant process optimization work and the results of the engineering studies aimed at design improvement. These improvements have resulted in a significant reduction in process costs.

Figures 8 and 9 show that the Advanced Coolside process enjoys a capital and levelized cost advantage relative to LSFO in all cases examined in this study. The figures further indicate that the economic targets established in the first interim evaluation have been achieved for a wide range of coal sulfur contents and plant sizes. The projected capital cost of Advanced Coolside is 55% to 60% lower than limestone forced oxidation wet FGD. The total levelized SO_2 control cost in \$/ton SO_2 removed ranged from 15% to 35% lower than LSFO, over the range of plant sizes and coal sulfur contents investigated. For a mid-range plant size (260 MW) and a mid-range coal sulfur content, the levelized cost advantage is about 25%. The levelized cost is sensitive to sorbent transportation charges and as a result is highly site-specific.

Using interim design and economic evaluations to provide direction for pilot plant optimization studies was instrumental in reducing the Advanced Coolside capital costs and levelized control costs. The capital cost was reduced by 30% compared with the interim evaluation completed in early 1993. Levelized SO_2 control costs were reduced by about 18% for a 260 MW plant. Much of the reduction resulted from reevaluating the equipment and process requirements in light of economic trade-offs. A discussion of the Advanced Coolside process improvements and resulting cost reductions is given in Appendix A. The major improvements include the use of the third generation contactor design (venturi contactor) for flue gas humidification and fly ash removal, optimization of the venturi contactor design and operating conditions, using hydroclones in place of thickeners for fly ash dewatering, and substituting a less-costly pugmill mixer for a high-intensity mixer for recycle solids water addition.

The Advanced Coolside conceptual process design is detailed in Appendix A. This appendix also gives the detailed assumptions used in the economic analysis. EPRI Technical Assessment Guidelines were followed. To achieve consistency for a comparative evaluation, similar design philosophies, equipment cost algorithms, and financial assumptions were used for the evaluation of both Advanced Coolside and limestone forced oxidation technologies. Both processes were evaluated for 90% SO_2 reduction. The process design for the limestone forced oxidation wet FGD process was recently updated based on current commercial trends to reflect the state of the art. This includes the use of a single absorber module with no spares.

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TABLE 1

SUMMARY OF RECYCLE TEST RESULTS FOR TESTS SIMULATING A BAGHOUSE PARTICULATE COLLECTOR,

Test	Fresh Ca/S, mol	Recycle Ratio (a)	lb Water per lb Recycle Sorbent	Total $Ca(OH)_2/S$ mol Ratio	Baghouse Approach, °F	SO_2 Removal, %		Sorbent Util., %	
						Duct	System (b)	Steady State (d)	Solids Analyses
12	1.4	4.5	0.15	2.2	23	83	90	63	62
13	1.2	6.9	0.12	2.1	23	87	90	75	70
12A (c)	1.5	4.3	0.15	2.5	24	84	90	60	59

Common Conditions: SO_2 Inlet Concentration = 1500 ppm (dry); Flue Gas Flow = 340 SCFM

(a) lb dry recycle/lb fresh lime
(b) duct + baghouse
(c) fresh lime and recycle sorbent wetted together and fed by one feeder
(d) calculated steady-state sorbent utilization

TABLE 2

SUMMARY OF RECYCLE TEST RESULTS FOR TESTS SIMULATING AN ESP PARTICULATE COLLECTOR

Test	Fresh Ca/S, mol	Recycle Ratio (a)	lb Water per lb Recycle Sorbent	Total $Ca(OH)_2/S$ mol Ratio	Baghouse Approach, °F	SO_2 Removal, %		Sorbent Util., %	
						Duct	System (b)	Steady State (d)	Solids Analyses
6A	1.2	5.0	0.00	2.2	10	59	73	61	58
7A	1.3	3.3	0.15	1.8	9	60	84	67	68
8A	1.2	3.4	0.10	1.8	11	64	81	65	66
9	1.5	3.5	0.15	2.2	12	70	90	61	63
10	1.2	4.9	0.15	1.7	9	81	88	71	68
11	1.6	3.9	0.15	2.4	11	91	97	60	58
11A	1.6	3.8	0.15	2.4	12	88	100	61	61
17B (c)	1.2	6.9	0.12	1.4	10	84	92	76	72

Common Conditions: SO_2 Inlet Concentration = 1500 ppm (dry)

(a) lb dry recycle/lb fresh lime
(b) duct + baghouse
(c) fresh lime and recycle sorbent wetted together and fed by one feeder
(d) calculated steady-state sorbent utilization

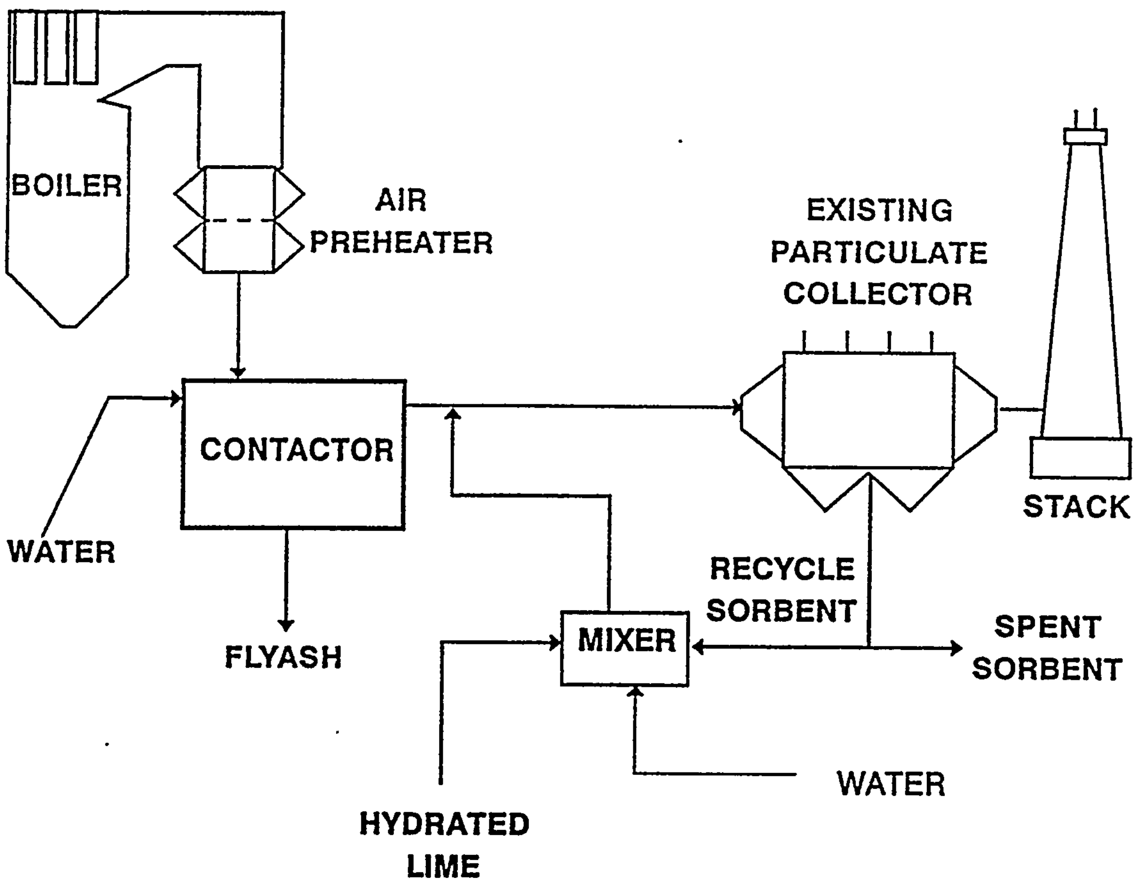

Figure 1. Schematic of the Advanced Coolside Process.

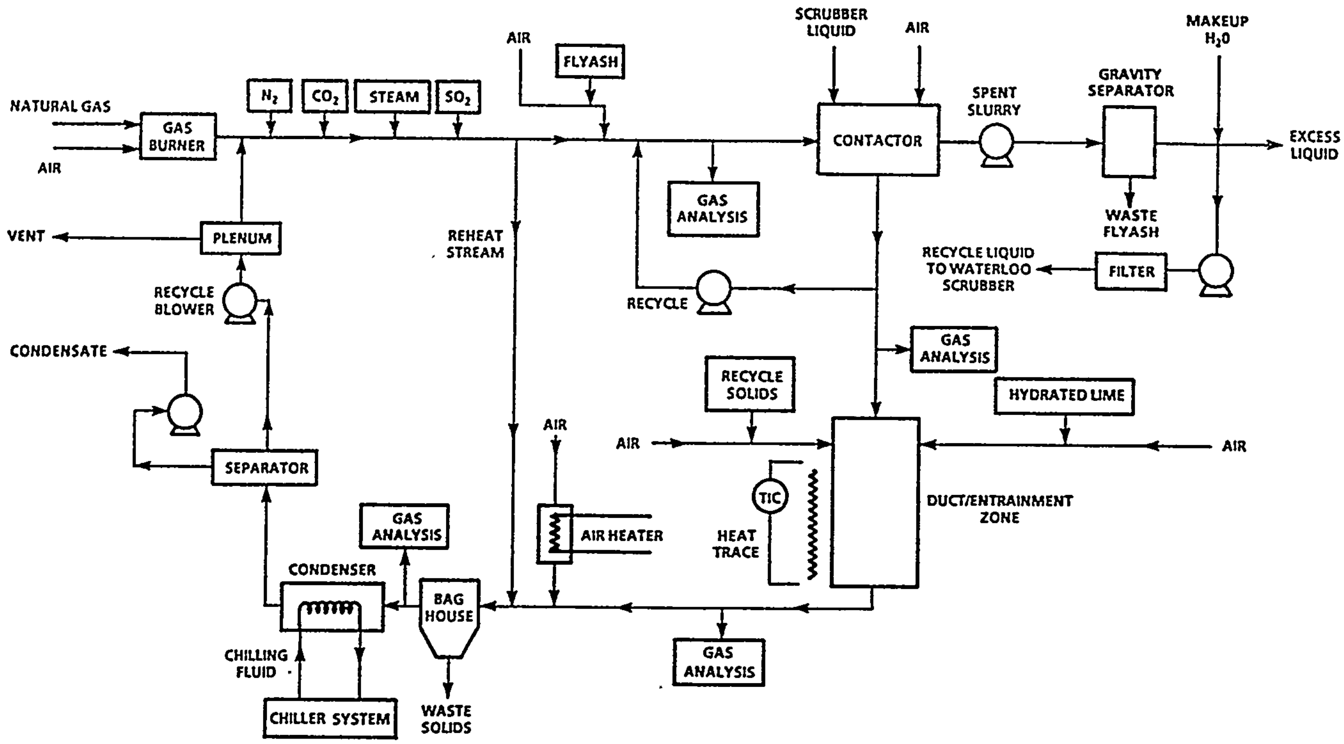

Figure 2. Schematic of the 1000 acfm Advanced Coolside Pilot Plant.

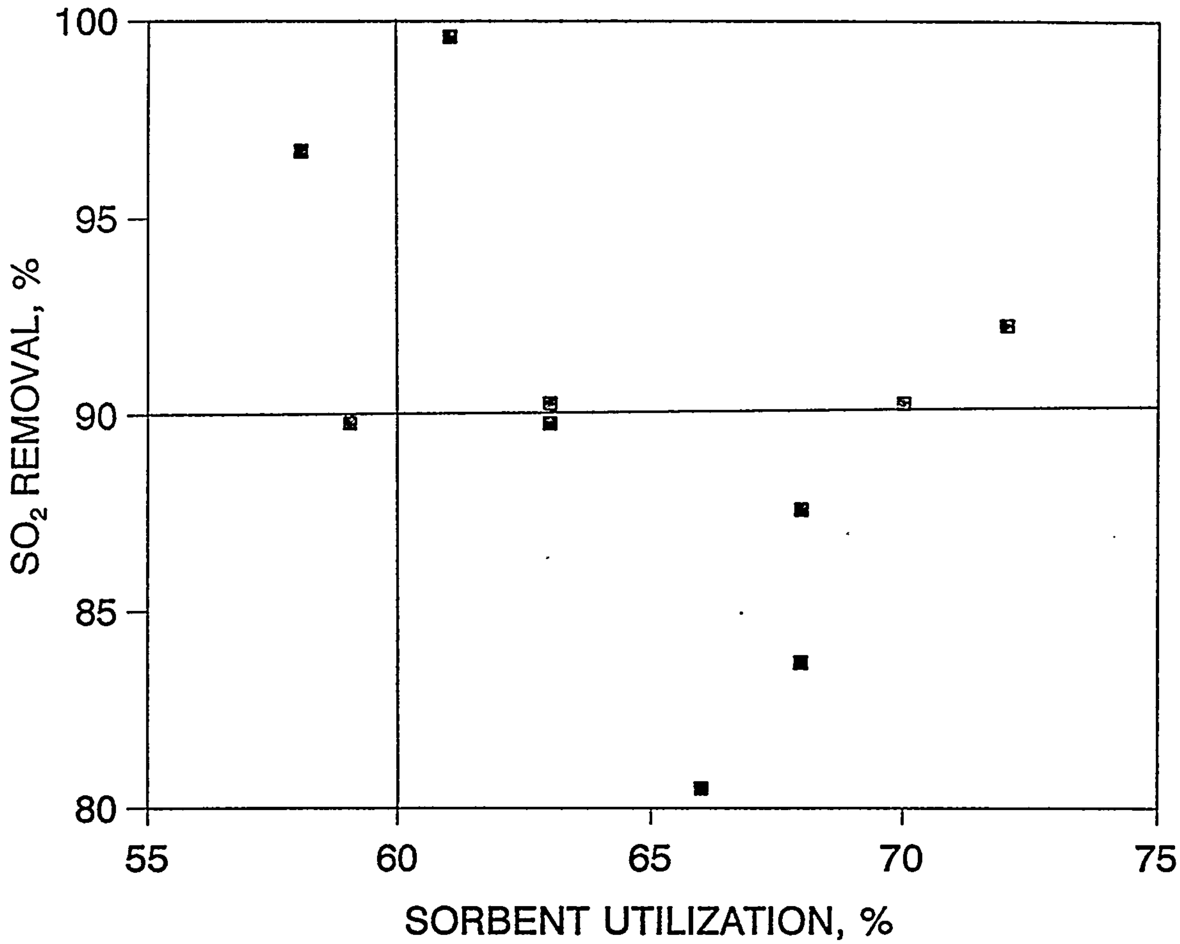

Figure 3. Recycle Simulation Test Results: System SO_2 Removals and Sorbent Utilizations.

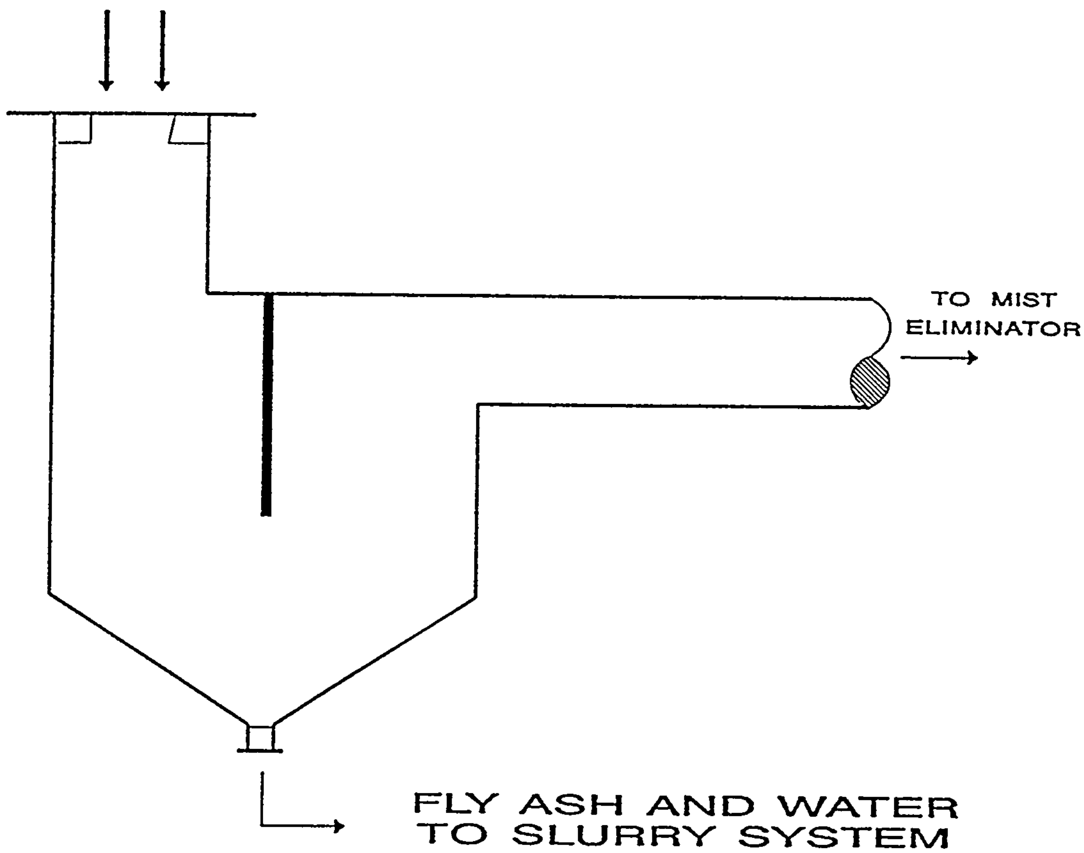

Figure 4. Schematic of the Second Generation Contactor

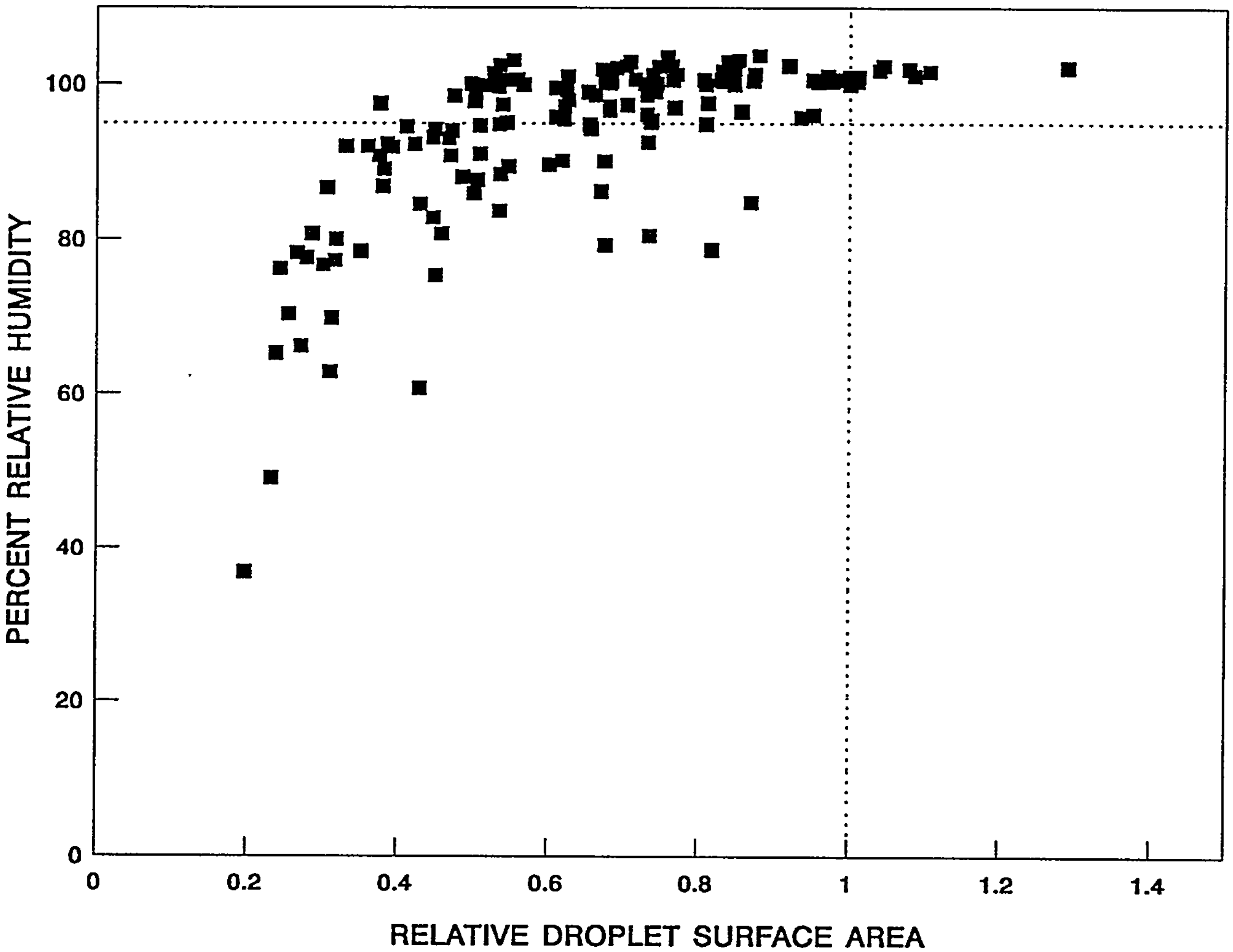

Figure 5. Saturation Efficiency for 150 Tests Using the Second Generation Contactor. Relative droplet surface area is the total droplet surface area (m^2/m^3 flue gas) divided by the design droplet surface area (per Turbotak, Inc.).

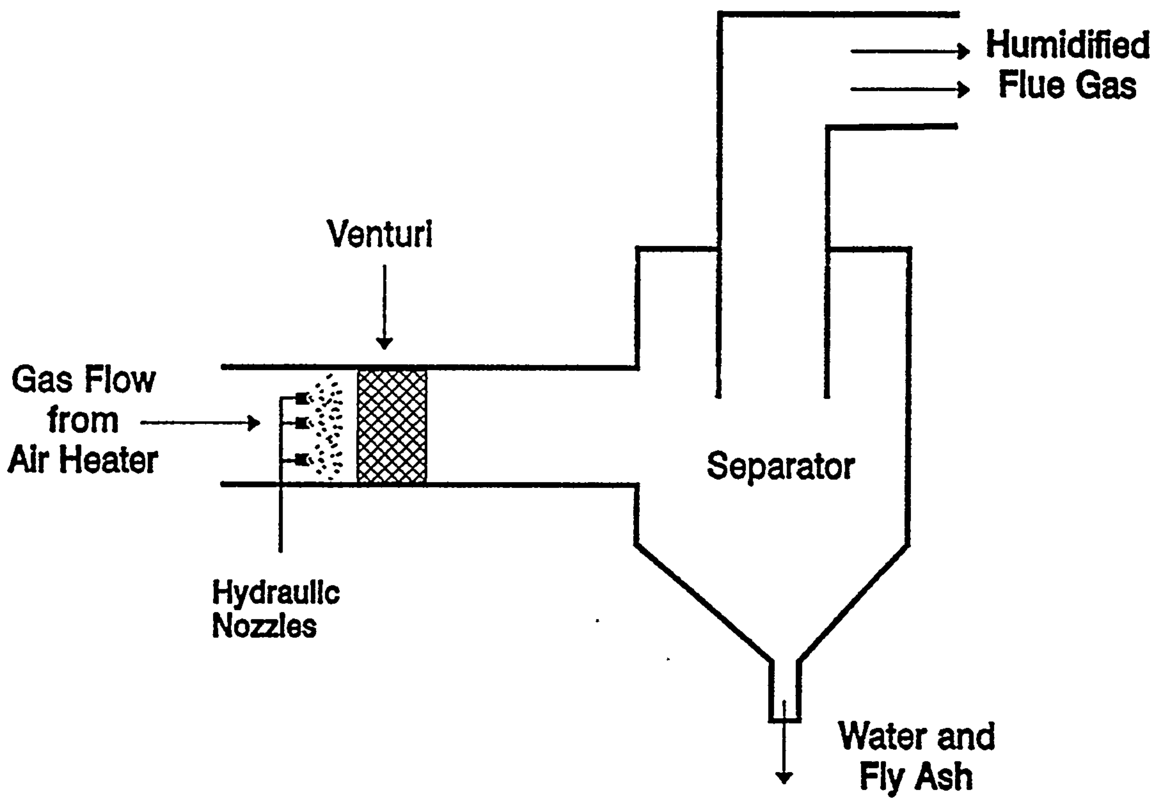

Figure 6. Schematic of the Third Generation Contactor.

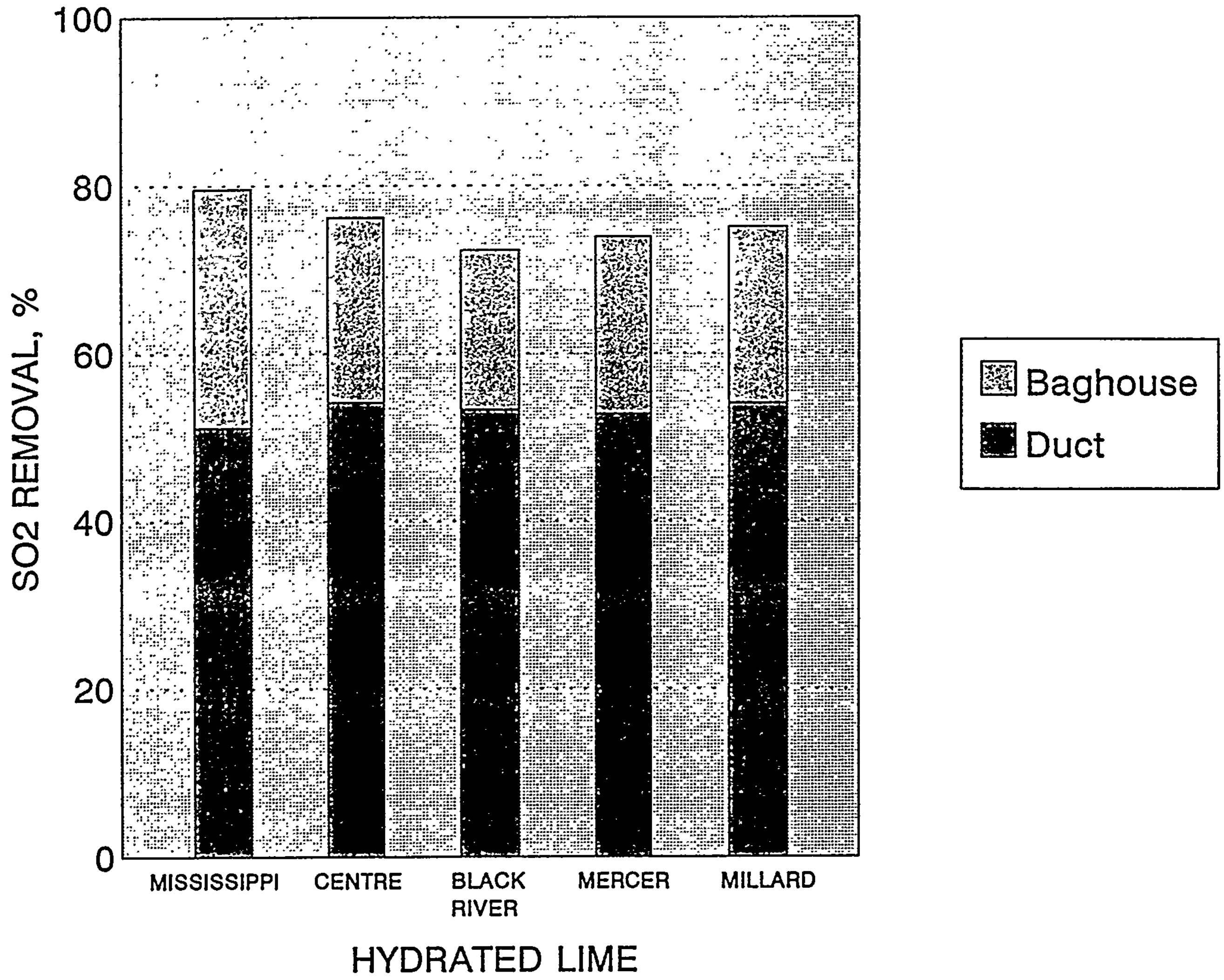

Figure 7. Duct and Baghouse SO_2 Removals Using Five Commercial Hydrated Limes at a 1.5 Ca/S Ratio.

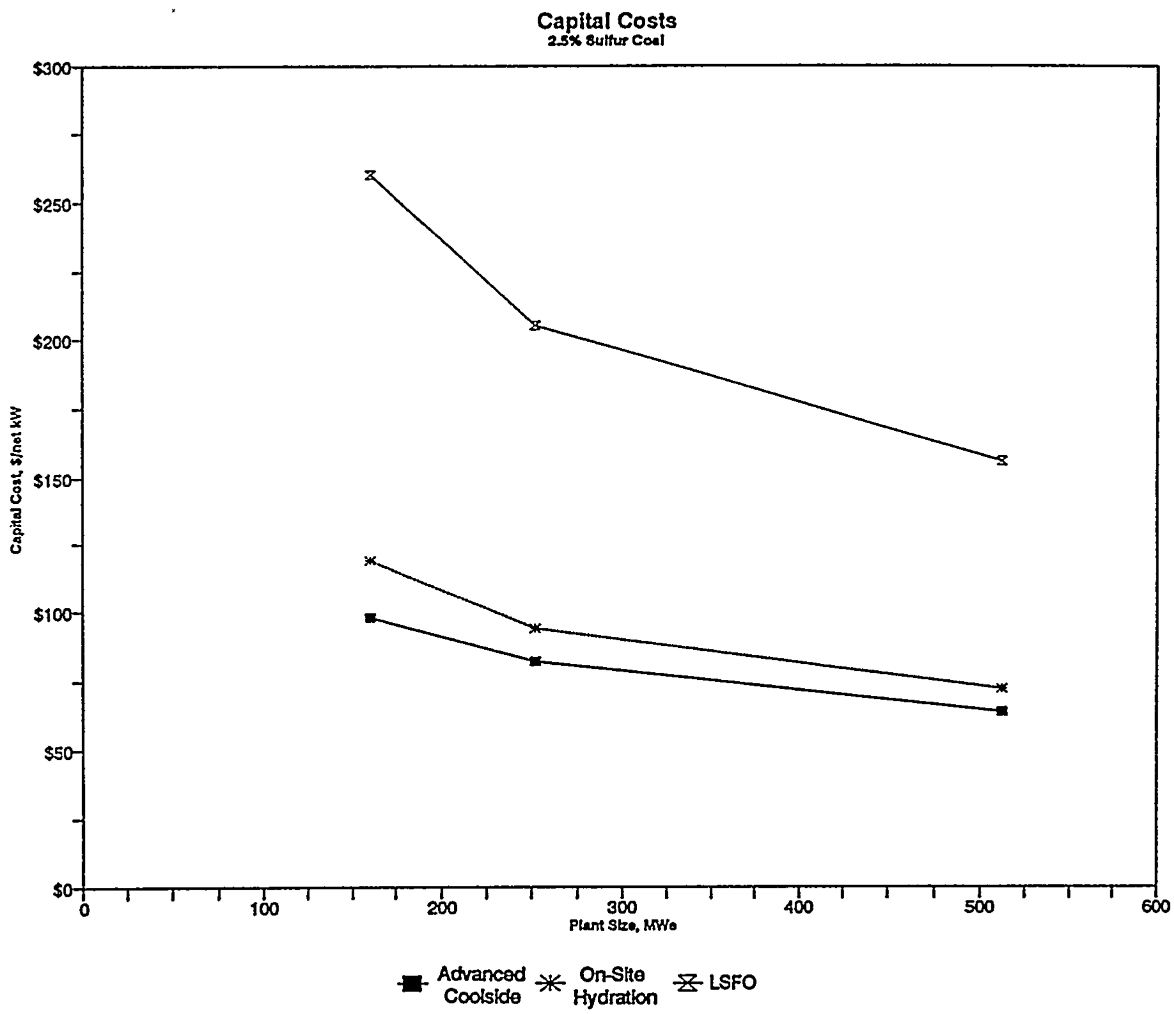

Figure 8. Comparison of Capital Costs for the Advanced Coolside Process and Limestone Forced Oxidation Wet FGD (LSFO) for a Range of Plant Sizes Burning a 2.5% S Coal.

(For Advanced Coolside, on-site hydration is compared to buying hydrated lime.)

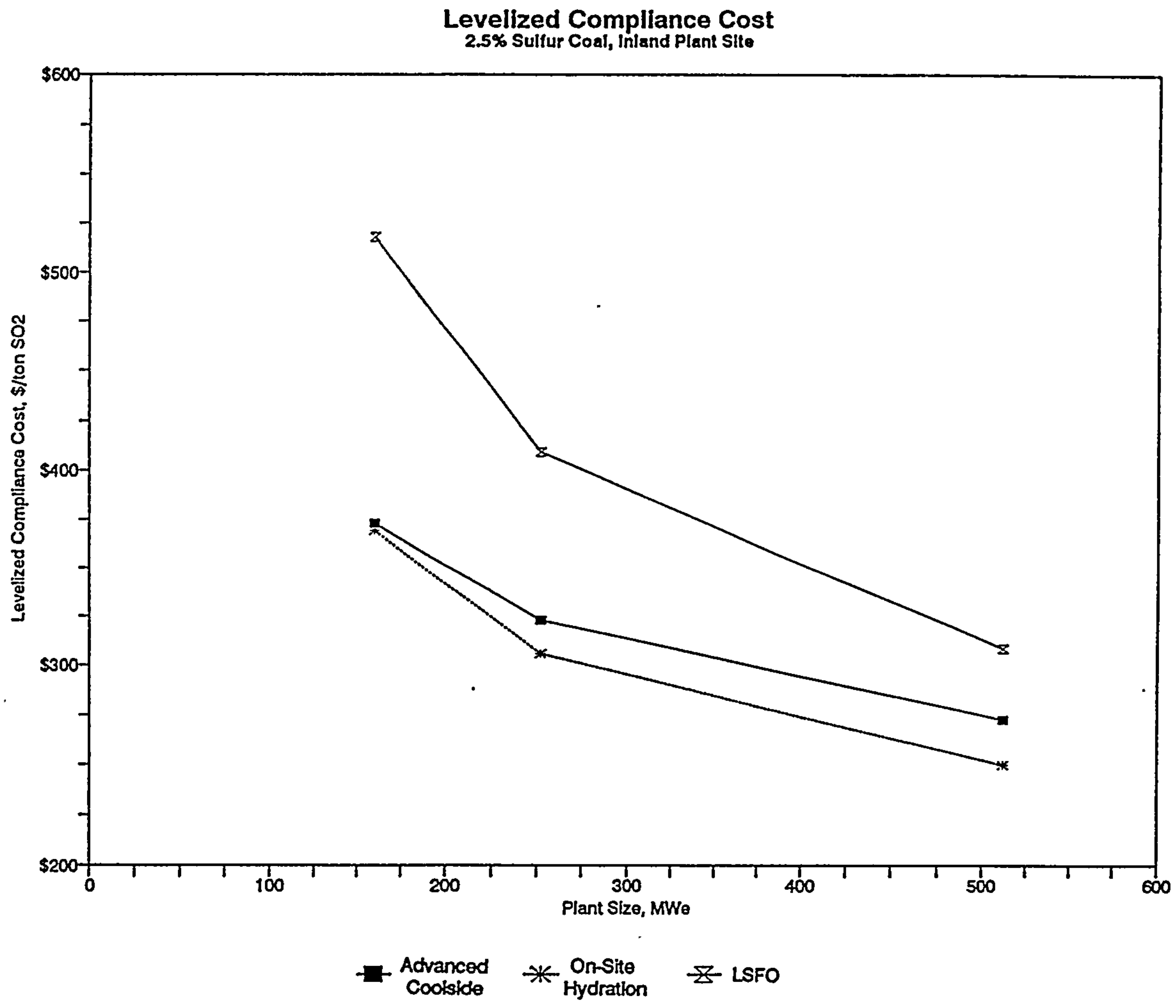

Figure 9. Comparison of Levelized SO_2 Control Costs for the Advanced Coolside Process and Limestone Forced Oxidation Wet FGD (LSFO) for a Range of Plant Sizes Burning a 2.5% S Coal and Assuming an Inland Plant Site.

(For Advanced Coolside, on-site hydration is compared to buying hydrated lime.)

APPENDIX A

CONCEPTUAL DESIGN AND ECONOMIC STUDY

Topical Report No. 6[18], discussing the conceptual design and economic evaluation, follows.

ADVANCED IN-DUCT SORBENT INJECTION FOR SO_2 CONTROL, DOE CONTRACT DE-AC22-91PC90360, TOPICAL REPORT NO. 6, TASK 5: CONCEPTUAL COMMERCIAL PROCESS DESIGN AND ECONOMIC EVALUATION

INTRODUCTION

The Advanced Coolside Desulfurization Process was developed through 1000 acfm pilot plant testing, as reported in Topical Report Nos. 1 through 5.[1-5] This development work showed the technical feasibility of the process and demonstrated that the original process performance targets could be exceeded. The 90% SO_2 removal target was achieved at sorbent utilizations up to approximately 75%, exceeding the target of 60% utilization. SO_2 removals in excess of 99% were achieved at utilizations greater than 60%.

The objectives of Task 6, Conceptual Design and Economic Evaluation, were to develop a conceptual design for a utility-scale application of the Advanced Coolside process and to assess the economic attractiveness of the process. Additional objectives of CONSOL were to identify process areas for potential cost reductions and to guide research and development efforts in areas that would most impact the economics and commercial readiness of the process. As a result, CONSOL began engineering and economic evaluation early in the project, and this was an ongoing process. Part of this evaluation by CONSOL involved the development of a heat and mass balance computer model which was used as a tool to help estimate process costs.

In early 1993, an interim process economic evaluation was completed. The interim study was initiated in order to explore the feasibility of an intermediate scale-up test of the process. Results indicated that Advanced Coolside had an economic advantage relative to limestone wet scrubbing for a range of plant sizes and coal sulfur levels. The evaluation identified several areas for potential process improvement, including equipment design optimization and sorbent utilization optimization. Areas identified for design optimization included improvement of the gas/liquid contactor design, improvement of the sorbent recycle handling system, and improvement of the waste handling system. As a result, it was decided to continue process optimization in the 1000 acfm pilot plant to explore these further areas of cost reduction. Pilot plant development work in these areas is described in Topical Report 2. Sorbent utilization optimization work is described in Topical Report 3.

Based on the results of the interim economic study, economic targets were established for the process. These were to achieve a 20% levelized cost advantage and a 50% capital cost advantage over limestone wet scrubbing for a range of plant sizes and coal sulfur levels.

In late 1993, CONSOL conducted a second interim process economics study. The study confirmed that projected SO_2 removal costs for the Advanced Coolside process were substantially reduced by the process design improvements established during pilot plant development work. In addition, the study showed that the cost advantage applied to a range of plant sizes and coal sulfur levels.

This report presents the results of a final process economic study for the Advanced Coolside process, under DOE Contract DE-AC22-91PC90360. It incorporates the results of recent pilot plant development work. It also includes results of the engineering studies aimed at design improvement.

The Advanced Coolside process was compared to the commercial technology of limestone forced oxidation (LSFO) for retrofit applications. The SO_2 abatement processes were evaluated at three plant sizes (160 MW, 262 MW, and 512 MW, gross) and four coal-sulfur levels (1.0%, 1.5%, 2.5%, and 3.5%, as-received).

The performance and economics of the technologies were assessed using the CONSOL Coal Quality Cost Model (CQCM), developed by CONSOL in the 1980s.[6] A process inlet flue gas flow rate and composition were estimated for each coal and plant size using the power plant module of the CQCM. These values were incorporated into an Advanced Coolside Cost Model (ACCM) and a LSFO model to provide the final process economics. The LSFO model was developed by CONSOL in the 1980s[6] and is regularly updated. Economic assumptions were based on EPRI technical assessment guidelines.

Capital costs for the two processes were compared and expressed as \$/net kW. In addition, detailed total compliance costs were determined for all scenarios investigated, in total levelized dollars and \$/ton SO_2 removed.

To achieve consistency for a comparative evaluation, similar design philosophies, equipment cost algorithms, and financial assumptions were used for the evaluation of both technologies.

CONCEPTUAL PROCESS DESIGN—ADVANCED COOLSIDE

The process flow for the Advanced Coolside process is categorized into fresh sorbent handling, sorbent preparation, flue gas flow, ash dewatering, and ESP waste handling sections.

FRESH SORBENT HANDLING

The hydrated lime handling area for the off-site hydration scenario is designed for rail delivery of hydrated lime. The hydrate is conveyed pneumatically from the railcars to the hydrate storage silo. The hydrate then is transferred from the storage silo to the duct injection point via the pneumatic injection blowers.

SORBENT PREPARATION

The pebble (quick) lime handling and preparation area for the on-site hydration scenario is similar to the off-site hydration area, except for the addition of hydrators. Pebble lime is pneumatically transferred from the unloading section to a day bin and hydrator feed bin. The pebble lime then is fed to the hydrator where water is added. The fresh hydrate is conveyed to the hydrate day bin while the grits, or insoluble residue, are fed to the grits bin. The hydrator is equipped with a vent scrubber and fan package for vent gas cleanup.

FLUE GAS FLOW

The flue gas flow area consists primarily of a venturi contactor, sorbent injection ports, and new duct run. It is assumed that the existing duct from the boiler splits into two trains each containing air heater and ESP modules.

To remove fly ash and humidify to saturation, the flue gas passes through the venturi contactor and contacts with coarse water sprays at the venturi throat. Pressure-drop-induced turbulence in the venturi throat breaks up the water droplets improving contact and vaporization. Total pressure drop across the venturi contactor is five inches of water. The water injection system in the venturi uses low-pressure, low-erosion nozzles. The system does not require a second fluid, such as air, and an associated compressor. Excess water and most of the fly ash are separated from the flue gas in the cyclone section of the venturi contactor and collect in the bottom. Once collected, the ash slurry is pumped to the dewatering section.

Prior commercial operating experience shows that the ESP can be successfully operated at an 18 °F approach to saturation. This study assumes that operation at a 10 °F minimum approach is possible; however, a reheat system is included in the design as a contingency. Like the return duct, the ESP is heat traced.

Once the flue gas passes through the ESP, it enters the existing ID fan and a new booster fan. A booster fan will not be required if the existing ID fan has sufficient excess capacity to cover the additional power requirement resulting from the Advanced Coolside process pressure drop. However, it is assumed that the existing ID fan is sized exactly for the existing (i.e., pre-retrofit) flue gas conditions. The booster fan is sized for the additional process pressure drop after correcting for the new process conditions. A steam reheater is included at the ID fan exit to assure sufficient stack buoyancy. It is designed to give a 30 °F approach to saturation.

<u>ESP WASTE/RECYCLE SOLIDS HANDLING</u>

Solids that are collected by the ESP are conveyed continuously from the ash hoppers to the recycle solids bin and the waste silo. Water is added to the recycle sorbent using a mixer. Once the water is added, the wetted sorbent is injected into the duct.

<u>ASH DEWATERING</u>

Dewatering of the venturi contactor bottoms is carried out with hydroclones. Use of hydroclones instead of a thickener results in a smaller footprint and lower capital cost.

Holding tanks are placed at the venturi contactor exit, hydroclone bank overflow, and hydroclone bank underflow. Pumps move the venturi contactor bottoms to the hydroclones and various other points in the process.

The fly ash and spent sorbent are disposed of by trucking to a land fill.

PROCESS DESCRIPTION—LIMESTONE FORCED OXIDATION

The limestone forced oxidation (LSFO) process is a standard post-ESP wet FGD process. The LSFO process uses the current state-of-the-art design for commercial operation. A single absorber module with no spare is assumed.

PROCESS DESIGN CONDITIONS

ADVANCED COOLSIDE

The Advanced Coolside process is assumed to operate at 90% total SO_2 removal and a fresh Ca/S ratio of 1.2, to yield a calcium utilization of 75%. SO_2 removal in the ESP is assumed to be 4% (absolute).

The pebble lime or hydrate storage silo has a capacity of 30 days while the silo feed blowers are sized for six times the required fresh lime feed rate. The recycle solids bin has a four-hour capacity.

For the on-site hydration scenarios, the commercially available hydrators are sized at either 10 or 15 tons per hour of product. One spare hydrator is supplied for each plant.

A pressure drop of 5" H_2O is estimated for the venturi contactor. Although the pilot plant venturi was operated at 6-8" H_2O, less pressure drop is expected in a commercial unit designed with a more gradual expansion after the throat. At these conditions, it is assumed that the venturi contactor removes 85% of the incoming fly ash and humidifies the flue gas to saturation. The contactor is designed to resist acid corrosion.

Corrosion-resistant material is used for the duct between the venturi contactor and the injection point. Since the presence of the alkaline solids eliminates acid corrosion, the new duct after solids injection is constructed of carbon steel.

The post-injection duct layout is configured to yield a total flue gas residence time of three seconds at 50 fps average velocity after lime injection. Half of the total residence time, or 1.5 seconds, is obtained in the new duct run while the remaining 1.5 seconds is obtained in the existing dual ducts. Process equipment layout considerations require much of this new duct length to provide reasonable access for maintenance. The total reaction duct requirement of three seconds is based upon engineering judgment of mixing conditions in the large ducts. The additional pressure drop resulting from the new duct run is estimated to be 1.5" H_2O.

Heat tracing of the ESP is included to insure that condensation does not occur on the walls. Ductwork from the venturi contactor through the ID fan is also heat traced. The electric costs correspond to operating the heat tracing at an annual average of 70% of design capacity.

Staffing of the Advanced Coolside process is set at an average of 3.25 operators per shift. This consists of three operators per shift, seven days a week, plus one operator on daylight during the five-day work week for waste disposal.

LIMESTONE FORCED OXIDATION

The LSFO Process is designed for 90% SO_2 removal and operates at a 1.05 available fresh Ca/S ratio. No additives are utilized in the system. A single absorber design philosophy is assumed for all plant sizes. Hydroclones are used for primary dewatering of absorber slurry. A new 350-ft high stack is assumed. Staffing for the LSFO Process is averaged at 4.2 operators/shift.

TECHNICAL AND ECONOMIC CRITERIA

The prices of consumables are listed in Table 1. Both lime and limestone prices are a function of site-specific delivery factors and may vary with location. A significant change in the delivered pebble lime or hydrate price will affect the economics of the technologies. For this report, the economics for generic delivered prices of pebble lime and hydrate for river (barge transport) and inland (barge plus rail/truck transport) locations were generated. Lime plant fob prices were set at $50/ton for pebble lime and $54/ton for hydrate. Barge transport rates were set at $4/ton for pebble lime and $5/ton for hydrate while truck/short rail rates were set at $3/ton and $6/ton, respectively. The difference in the transport rates for pebble lime and hydrate reflect truck/car capacities for the different bulk densities (60 lb/cf for pebble lime versus 35 lb/cf for hydrate).

Specifications for the 2.5% sulfur coal are listed in Table 2. The coal represents a cleaned, eastern bituminous product.

Design assumptions for the processes are 90% SO_2 removal, 65% net capacity factor, and 30-year capital life. Indirect costs, expressed as a percentage of direct costs, consist of 13.8% field costs, 22.4% home office, and 1% bonds, all-risk insurance, and tax. An 18% contingency is used for all technologies.

A medium-difficulty retrofit level and a standard 1.06 location factor are assumed for all technologies. A two-year construction life is used for Advanced Coolside while LSFO is based on a three-year construction life. Other assumptions are a 4.5% inflation rate, 45% debt, and a 38% income tax rate. All costs are in 1992 dollars.

ECONOMIC RESULTS

Predicted capital costs and total annual levelized costs for the Advanced Coolside process are listed in Tables 3 and 4. The capital costs are expressed in \$/net kW of capacity while the total annual levelized costs are expressed in \$/ton SO_2 removed. Note that these costs do not include coal or other boiler-related expenses. As a result, the costs in Table 3 represent the total additional SO_2 control cost that results from the capital expenditure costs, operating costs, maintenance costs, and variable costs attributed solely to the FGD process. Plots comparing the capital and levelized compliance costs for Advanced Coolside and wet FGD (LSFO) for the 262 MW plant cases are shown as Figures 1 and 2.

Advanced Coolside has significantly lower capital cost requirement than LSFO for all cases investigated. The capital cost advantage for the Advanced Coolside process over LSFO ranges from 50% for the 1.0% sulfur coal, 160 MW plant, on-site hydration case up to 62% for the 2.5% sulfur, 160 MW plant, off-site hydration case. For the 512 MW plant cases, the capital cost advantage ranges from 53% for the 1.0% sulfur coal, on-site hydration case up to 59% for the 3.5% sulfur coal, off-site hydration case.

Advanced Coolside enjoys a total levelized cost advantage relative to LSFO for all cases. The total removal cost advantage for the Advanced Coolside process, at a river location, relative to LSFO, on a \$/ton SO_2 removed basis, ranges from 17% for the 3.5% sulfur coal, 512 MW plant, on-site hydration case to 35% for the 1.0% sulfur, 160 MW plant, off-site hydration case. For the 262 MW plant burning a 2.5% sulfur coal and employing on-site hydration, the Advanced Coolside compliance cost advantage is 26%.

For the 262 MW plant burning a 2.5% sulfur coal, adding the hydrator to the Advanced Coolside process increases the required capital by \$12/kW but decreases the overall removal cost by \$12-17/ton of SO_2 removed, depending on the reagent delivered prices.

PROCESS IMPROVEMENTS

A number of key process improvements have been added to the Advanced Coolside process since the initial interim economic study. For the 262 MW plant size burning a 2.5% sulfur coal, total process capital was reduced by approximately \$6.8 MM, which translates to over \$62/ton SO_2 removed. Since these are capital cost savings, the levelized cost, \$/ton SO_2 removed, is much higher for the low-sulfur coals. For the 1% sulfur coal, the savings is 2.5 times the previously mentioned \$62/ton SO_2 removed. This reduction was a result primarily of switching to a venturi contactor for fly ash removal and humidification (~\$4.0 MM), using hydroclones in place of a thickener for ash dewatering (~\$2.0 MM), and improving the recycle handling system (~\$0.8 MM). The process improvements were corroborated by either pilot plant tests or by engineering studies and vendor recommendations.[2]

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TABLE 1

PRICES

	Price
Pebble Lime, River/Inland	$54/$60/ton
Hydrate Lime, River/Inland	$57/$65/ton
Limestone	$15/ton
Water	$0.60/Mgal
Fly Ash Credit	$8/ton
FGD Waste Disposal	$6.50/ton
Replacement Power	$30/MW
Operating Labor	$22/hr
Maintenance Labor	$18.90/hr
Administration	$16.87/hr

TABLE 2

COAL SPECIFICATIONS

Coal Sulfur Level	2.5% S
<u>Proximate Analysis, wt %</u>	
Moisture	5.5
Volatile Matter	36.5
Ash	7.5
Sulfur	2.5
Heating Value, Btu/lb	13,200
<u>Ultimate Analysis, wt % dry</u>	
Hydrogen	5.2
Carbon	77.5
Nitrogen	1.4
Oxygen	5.2
Sulfur	2.7
Ash	7.9
Chlorine	0.1
Heating Value, Btu/lb	13,968

TABLE 3

SUMMARY OF COST

Coal Sulfur	Plant Size	Pebble & Hydrate	Capital Cost		Levelized Cost	
			Advanced Coolside w/o Hydrator	Advanced Coolside w/Hydrator	Advanced Coolside w/o Hydrator	Advanced Coolside w/Hydrator
% AR	MW	$/ton	$/Net KW	$/Net KW	$/ton SO_2	$/ton SO_2
River Site						
2.5	262	54/60	82	94	315	303
Inland Site						
2.5	262	57/65	82	94	323	306

TABLE 4

DETAILED COSTS OF 262 MW, 2.5% SULFUR COAL CASE
FOR RIVER DELIVERY

Process	Advanced Coolside	
Hydration	Off-site, $	On-Site, %
Capital Section		
Reagent Preparation	2.103	3.552
Sorbent Injection	0.807	1.102
Venturi Train	1.554	1.554
Flue Gas Handling	4.265	4.265
Reaction Duct/Absorber	0.166	0.166
Recycle System	0.897	0.897
Particulate Collection	0.215	0.215
Reheat	0.248	0.248
Waste Handling	1.665	1.665
Chimney	0.000	0.000
Miscellaneous	0.715	0.820
Total Direct	12.635	14.484
Field	1.744	1.999
Home Office	2.831	3.244
Bond, ARI, Tax	0.126	0.145
Contingency	3.121	3.577
TPI	20.457	23.449
$/net KW	82	94
Levelized Cost Section		
Capital		
Levelized TPI	2.117	2.427
Preproduction	0.195	0.205
Working Capital	0.136	0.138
Total Capital	2.448	2.770
Variable O&M		
Reagent	2.235	1.547
Water	0.054	0.054
Waste Disposal	0.543	0.543
Power	0.537	0.552
Total Variable O&M	3.369	2.696
Fixed O&M		
Operating Labor	0.626	0.626
Maintenance	0.522	0.590
Administration	0.250	0.259
Total Fixed	1.398	1.475
Total O&M	4.766	4.171
Total Levelized Cost	7.214	6.941
$/ton SO_2 Removed	315	303

*Note: Costs are expressed in $MM unless stated otherwise.

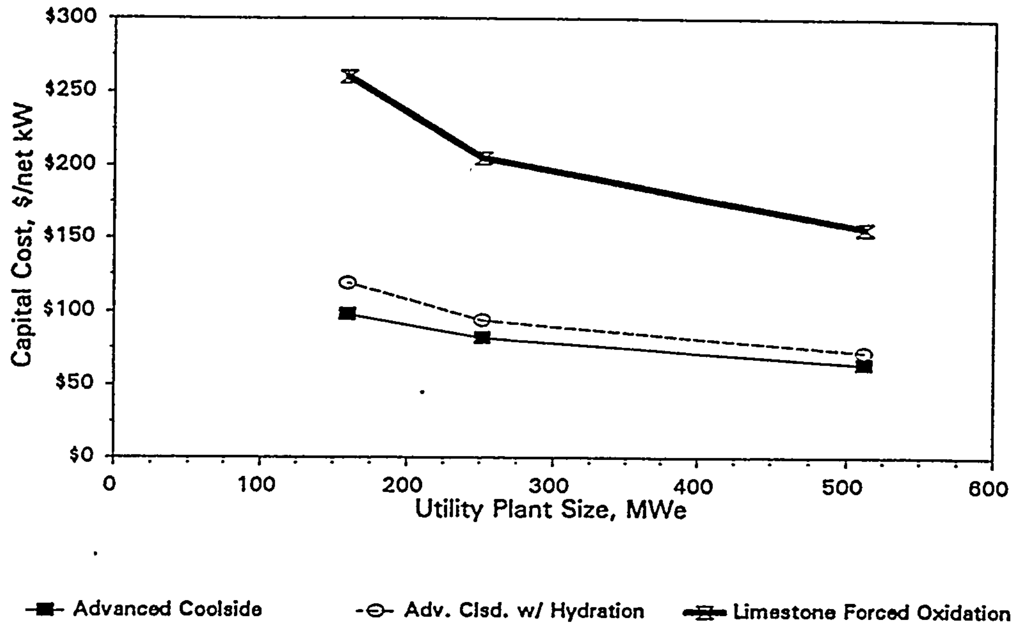

Figure 1. FGD Capital Costs (2.5% Sulfur Coal).

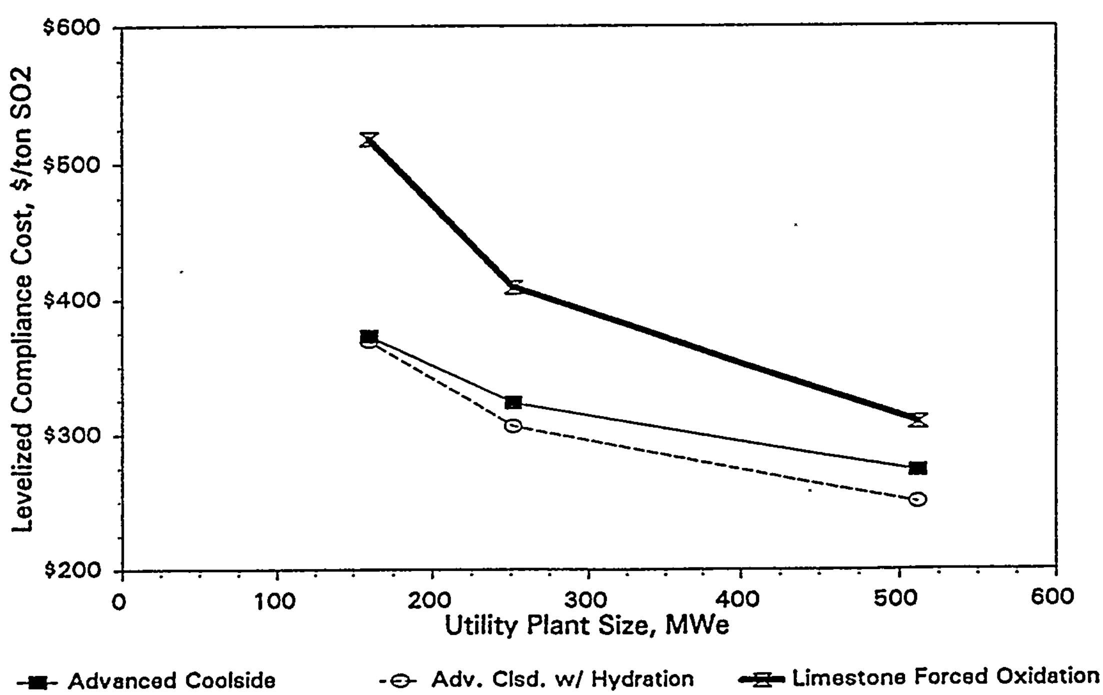

Figure 2. Levelized Compliance Cost for FGD (2.5% Sulfur Coal, Inland Plant Site).

APPENDIX B

DESIGN OPTIMIZATION

An excerpt of Topical Report No. 2,[14] which discusses recycle optimization and equipment design optimization, follows. The Conclusions, Experimental, and Discussion sections, figures, and selected tables are included.

CONCLUSIONS

RECYCLE OPTIMIZATION

1. By optimizing recycle, 90% SO_2 removal was achieved at sorbent utilizations of up to 75%, exceeding the original performance target of 60% sorbent utilization. At a 1.2 fresh Ca/S ratio, a recycle ratio of 7 lb recycle/lb lime, and a water addition level of 0.12 lb/lb recycle, SO_2 removal was 87% in the duct and 91% across the system. In this test, the reaction in the baghouse was partially quenched to simulate the SO_2 removal expected in an ESP.

2. Recycle tests showed that the Advanced Coolside process has the potential for very high SO_2 removal. With a baghouse, SO_2 removal greater than 99% was achieved at a sorbent utilization efficiency exceeding 60%.

PROCESS EQUIPMENT DESIGN OPTIMIZATION

1. A second generation contactor, consisting of a spray zone and a mist eliminator, showed good performance in pilot plant tests in terms of humidification efficiency, particulate collection efficiency, and operability. Downstream desulfurization results were equivalent to those with the Waterloo scrubber; the recycle results reported above were obtained using the second generation contactor. Operating conditions, such as water flow and atomization conditions, were optimized through parametric tests in order to reduce operating and capital costs.

2. Pilot plant tests indicated the feasibility of a third generation contactor design, consisting of a low-pressure-drop venturi and a cyclonic separator. The particulate collection efficiency and operability were good. The level of humidification achieved was typically within 1 to 4 °F approach to saturation. Downstream desulfurization performance was slightly less than with the second generation contactor design; 5 to 15% higher Ca/S ratio was required to achieve the performance target. The injection of a small amount of steam at the cyclonic separator exit was found to give a closer approach to saturation (0 to 1 °F) and to increase desulfurization performance. With steam injection, the required Ca/S feed rate with the venturi contactor was about the same as with the second generation design.

3. A pugmill was shown to be effective for recycle moisture addition, in place of the high intensity mixer previously used. Pilot tests were conducted in conjunction with a commercial vendor. Operability was good. The recycle material wetted in the pugmill showed the same desulfurization activity as that wetted in the high intensity mixer. The tests provided cost and scale up data.

DESCRIPTION OF ADVANCED COOLSIDE PROCESS

Figure 1 shows a schematic of the Advanced Coolside process. The process achieves greater SO_2 removal and sorbent utilization than previous duct sorbent injection processes by operating at a higher flue gas humidity and by more fully exploiting the potential of sorbent recycle. The key to the process is a gas/liquid contacting device downstream of the air preheater. The contactor serves two purposes: to nearly saturate the flue gas with water and to remove most of the coal fly ash from the flue gas. The sorbent is injected downstream of the contactor into the highly humid flue gas. Hydrated lime is very active for SO_2 capture near the saturation point, even in the absence of liquid water droplets. Because the flue gas is already humidified prior to sorbent injection, there is no strict residence time requirement for droplet evaporation. SO_2 is removed by the sorbent in the duct and by that collected in the existing electrostatic precipitator (ESP) or baghouse. The heat of reaction between SO_2 and hydrated lime raises the temperature of the flue gas by roughly 8-10 °F for each 1000 ppm of SO_2 removed. Therefore, the particulate collector can be operated at an elevated approach to saturation without flue gas reheat. However, because hydrated lime activity is highly sensitive to the approach to saturation, this reaction heat effect also acts as a limiting mechanism for SO_2 capture.

The spent sorbent is captured by the existing particulate collector as a dry powder. Sorbent recycle is an integral component of the Advanced Coolside process. Laboratory and pilot plant tests have shown that recycle sorbent is quite active for SO_2 capture at high humidity. The potential for recycle is increased because fly ash is removed separately before sorbent injection. Furthermore, process performance can be improved by adding small amounts of H_2O to the recycle sorbent prior to re-injection. The water acts to maintain a close approach to saturation by evaporating, thus, counteracting the heat of reaction.

Equipment design optimization has focused on the flue gas/water contactor. For the initial pilot plant testing the contacting device was a Waterloo scrubber.[14,15] This is a commercially available device, marketed by Turbotak Technologies Inc. Design and testing of improved contactor designs, substantially simpler than the Waterloo scrubber, was one of the key work areas described in this report.

EXPERIMENTAL

PILOT PLANT DESCRIPTION

Figure 2 is a schematic of the Advanced Coolside desulfurization pilot plant. It was designed to simulate integrated Advanced Coolside operation with combined flue gas saturation and fly ash removal using a contactor, and sorbent injection downstream of the contactor into the saturated flue gas. The plant consists of a flue gas generation system, a flue gas/water contactor, a spent slurry handling system, sorbent injection systems, a recycle sorbent moisture addition system, the test duct section/reactor, a baghouse, a flue gas cooling and recycle system, and flue gas analyses systems.

Flue Gas Generation System

A simulated flue gas stream is produced by mixing the combustion products of a natural gas combustor, bottled gases (SO_2 and CO_2), steam, plant N_2, and fly ash with a recycle gas stream from the process. Gas flows of 150 to 350 scfm are generated in this manner. The gas combustor serves to generate make-up flue gas and to raise the flue gas temperature to the desired level. The inlet SO_2 content can be varied from 500 to 2500 ppm by SO_2 injection. The fly ash loading can be varied from 0 to 5 gr/scf. The flue gas adiabatic saturation temperature at the baghouse exit is controlled by the rate of steam injection upstream of the contactor.

Recycle of the flue gas provides over 80% of the total flow. After particulate removal, the flue gas is cooled by a heat exchanger, and condensed water is removed by an impaction separator. The gas then is recycled by means of a blower.

Contactor

Three different contactor designs were evaluated as part of this test program. These included a Waterloo scrubber provided by Turbotak Inc., a second generation design consisting of a spray zone followed by a mist eliminator, and a third generation design consisting of a venturi followed by a centrifugal separator. The designs of these contactors are discussed in detail later in this report. All the contactors were designed to remove fly ash and nearly saturate the flue gas with water.

To increase contactor gas throughput, a contactor recycle fan provides the ability to recirculate flue gas from the contactor exit directly back to the contactor inlet; this was provided because the original Waterloo scrubber was designed for 1000 acfm, which is higher than the maximum output of the flue gas generation system. During operation of this contactor recycle fan, coal-fired post-air-preheater flue gas conditions are simulated by mixing high-temperature (~450 °F) flue gas from the flue gas generation system with the low-temperature flue gas recycled from the contactor exit. When the contactor recycle fan is not in use (i.e., low contactor throughput tests), the flue gas generation system is regulated to supply lower temperature (ca. 300 °F) flue gas to the contactor inlet.

Spent Slurry Handling

Solids are removed as a slurry from the contactor. The spent slurry is pumped to a gravity separator. Solids are removed as a sludge (approximately 50/50 fly ash/water by weight). The clarified liquor is recycled to the contactor. Make-up fresh water is added as needed.

Flue Gas Ductwork

The saturated flue gas exiting the contactor passes through 53 ft of 5-inch pipe, then 24 ft of 8-inch pipe and then 7 ft of 4-inch pipe before entering the baghouse. This simulates the ductwork in the Advanced Coolside process, providing contact between sorbent particles and the humid flue gas. Sorbent was injected into the 5-inch pipe near the contactor exit.

Recycle Sorbent Moisture Addition

A batch mixer manufactured by Littleford Bros., Inc. is used for tests with water addition to recycle sorbent.

Baghouse

A baghouse is used to remove particulates from the duct effluent flue gas. It is a pulse-jet type baghouse with 9 bags, giving a total cloth area of 144 ft^2. Solids are collected in a 55-gal drum under the baghouse hopper. The hopper is sealed by a butterfly valve. The baghouse is heat traced to maintain adiabatic operation.

The flue gas can be reheated before the baghouse to control the approach to saturation at the baghouse exit. Reheat is accomplished by injecting hot air between the duct exit and the baghouse. The approach to saturation at the baghouse exit can be varied from 10 to 25 °F. As mentioned above, a 10 °F approach can be maintained without reheat, due to the reaction heat effect.

Flue Gas Analysis

The flue gas composition is measured continuously by on-line analyzers at four locations: the contactor inlet, the contactor exit (duct inlet), the duct outlet, and the baghouse exit. This allows measurement of SO_2 removals in the contactor, in the ductwork, and in the baghouse. SO_2 and O_2 contents are measured at all locations. The O_2 content is used to correct for air in-leakage. Gas sampling systems are designed to prevent further reaction of SO_2 with the sorbent particles. The flue gas analyzers are also used to control the concentration of SO_2, N_2 and CO_2 into the system.

TEST PROCEDURES

Recycle Optimization Tests

Simulation of Steady-State Recycle. The Advanced Coolside pilot plant can be operated to closely simulate continuous, steady-state sorbent recycle. Spent sorbent from the baghouse is removed periodically during operation. A portion is returned to the recycle feeder hopper with or without moisture addition, and a portion is discarded. The recycle then is fed continuously back into the flue gas simultaneously with fresh hydrated lime. A recycle simulation test is initiated using baghouse solids from previous once-through or recycle tests as the recycle sorbent.

In some of the previously reported recycle simulation tests, the conditions for steady-state continuous recycle were approached, but not fully established.[13] The sorbent utilizations based on analyses of baghouse solids were lower than the calculated utilizations based on gas analyzer and the sorbent feed rate data. In the test program reported here, steady-state conditions were more closely approached by extending the duration of the tests and by reducing the inventory of sorbent in the system. Each test was continued until on-the-spot solid analyses confirmed that steady-state recycle conditions had been established, or closely approached.

Simulation of ESP and Baghouse Performance. In the recycle tests there were two modes of baghouse operation. In some tests, the baghouse exit was operated at a relatively high approach to adiabatic saturation (20-25 °F) by reheating the gas entering the baghouse with hot air injection. This was done to limit the SO_2 removal in the baghouse and roughly simulate the SO_2 removal expected in an ESP. Previous experience with the Coolside process[1,5,6] indicates that ESP removal is limited by gas phase mass transfer; because of the high sorbent activity, this is expected to be the case for the Advanced Coolside process as well.

In other tests, no reheat was used and the baghouse exit was operated at a lower approach (10-13 °F) to maximize the SO_2 removal in the system. In the later case, the heat of reaction in the duct and the baghouse was sufficient to maintain the 10-13 °F approach.

Moisture Addition. In most of the tests reported here, the recycle sorbent was wetted using a high-intensity batch mixer, which is described above. Procedures for wetting recycle were selected to optimize operability, based on prior CONSOL experience.

One batch of sorbent was treated in a continuous, pilot-scale (100-1200 lb/hr) pugmill at the test facilities of Heyl & Patterson, Inc. This material was tested in the pilot plant to compare its desulfurization performance with that of sorbent prepared in the batch mixer. These results are discussed in detail in a later section.

Sorbent Injection. In all the recycle tests reported here, the sorbents (fresh and recycle) were injected into the humidified duct at the contactor exit. Between the sorbent injection point and the baghouse was 84 ft of ductwork, giving an in-duct gas residence time of 2.7 sec. A more detailed description of the ductwork is given in the section on operability.

Performance Measurement. The SO_2 removals reported are based on the readings of the continuous gas analyzers located at the contactor exit prior to sorbent injection, the duct exit, and the baghouse exit. The removals reported were measured near the end of a test, after steady-state recycle conditions had been established or closely approached. The data represent an average over time periods during which the process conditions (temperatures, sorbent feedrates, flue gas flow, etc.) were lined out.

The system (duct + baghouse) SO_2 removals were confirmed in each test by analysis of the baghouse solids. In one test, the in-duct SO_2 removal was confirmed by EPA Method-6.

<u>Contactor Operation</u>. In all the recycle optimization tests, the contactor was operated to achieve near saturation conditions. Two different contactors were used: the Waterloo scrubber and the second generation contactor consisting of a spray chamber and mist eliminator. The second generation contactor was used in the majority of the tests. Since saturation was closely approached in all tests, the contactor used did not affect desulfurization performance. Operating conditions for achieving saturation are given in the discussion of contactor optimization later in this report.

The flue gas temperature at the contactor inlet was controlled at ~300 °F by controlling load on the gas combustor. The flue gas adiabatic saturation temperature was controlled at ~125 °F at the baghouse exit by controlling the rate of steam injection upstream of the contactor. The saturation temperature at the control point (baghouse exit) was lower than that at the contactor exit or that at the duct exit due to dilution by sorbent transport air and by flue gas reheat air. A rough calculation based on estimated dilution flows indicated that the saturation temperature was ~3 °F higher than the control point at the contactor exit and ~1 °F higher at the reactor/duct exit during once-through operation.

No fly ash was injected into contactor inlet flue gas for any of the recycle optimization tests. This was done for test simplicity. As discussed in Topical Report No. 2, tests confirmed that the contactors removed over 95% of the inlet fly ash and that the presence of fly ash in the inlet gas did not affect downstream desulfurization performance.

<u>Contactor Optimization Tests</u>

<u>Contactor Operation</u>. The contactor recycle system was used during this test program to enable the contactor to handle higher gas throughput than provided by

the existing flue gas generation system. This consists of a fan which recycles some of the flue gas from the contactor exit to the inlet. Flue gas flows of 400 to 1000 acfm at 270 to 290 °F were generated by controlling recycle flow and the temperature of the gas from the flue gas generation system.

Fly ash was injected into the flue gas upstream of the contactor only for the particulate collection efficiency tests; for test simplicity, fly ash was not injected during other tests.

Details of operation for each contactor design tested are provided in a later section.

Flue Gas Duct Operation. A minimal amount of heat was applied to the duct walls to simulate near-adiabatic operation of a large duct.

Baghouse Operation. The normal operation of the baghouse was changed because of the difficulty in measuring relative humidity near the saturation point. The flue gas relative humidity is normally determined from a psychometric chart using the measured wet bulb and dry bulb temperatures. However, at conditions very close to saturation, there is little driving force for evaporation from a wetted wick thermocouple and, consequently, the wet bulb temperature measurement can be less reliable. In the saturation efficiency tests there was no sorbent fed to the duct and, thus, no exothermic heat of reaction between SO_2 and sorbent to raise the gas temperature. As a result, the flue gas temperature was nearly equal to the wet bulb temperature when the gas exited the duct/reactor. Electrically heated plant air (300 to 350 °F) was mixed with the flue gas at the duct exit upstream of the baghouse to raise the flue gas temperature to about 25 °F above the wet bulb temperature; wet bulb temperatures are more accurately measured under these conditions. The amount of added air was calculated based on the reheat gas temperature and the flue gas temperature measured at the baghouse inlet and exit; this number then was used to calculate the relative humidity of the undiluted, unheated flue gas.

During the SO_2 removal tests the heated plant air was used in the same manner to increase the baghouse flue gas temperature for tests made at 18 to 25 °F approach to saturation. Tests made at a 10 to 15 °F approach required little or no reheat due to the exothermic heat of reaction.

Fly Ash Feed. The fly ash used during the particulate collection efficiency tests came from three sources: the Cleveland Electric Illuminating Company Avon Lake Power Plant, the Duquesne Light Company Elrama Power Plant, and New York State Electric and Gas Company Kintigh Power Plant. All three were bituminous coal fly ashes with mass mean diameters in the range 10 to 15 μm.

Fly Ash Collection Efficiency Measurement. Fly ash grain loading was measured at the contactor exit using EPA Method 17. The inlet fly ash loading was determined based on the weighed fly ash feed and the flue gas flow determined by pitot tube traverse.

Water Droplet Size and Surface Area. The water droplet Sauter mean diameters produced by the water spray nozzles were calculated from the air pressure and water flow rate using information supplied by the contactor supplier. The total droplet surface area was calculated from the Sauter mean diameter and the total water flow.

TEST PROGRAM

Recycle Optimization Tests

In the recycle optimization tests, the following variables were investigated over the indicated ranges:

- Fresh Ca/S mol ratio: 1.2 - 1.6
- Recycle Ratio (lb dry recycle/lb fresh lime): 3.3 - 6.9
- Moisture Addition to Recycle Sorbent (lb water/lb recycle sorbent): 0.00 - 0.15
- Approach to Adiabatic Saturation at the Baghouse: 9 - 24 °F
- In-duct Gas Residence Time: 1.0 - 2.7 sec

For each test the inlet SO_2 concentration was 1500 ppm (dry), and the fresh sorbent was Mississippi hydrated lime (see Table 1). The test conditions and results are summarized in Tables 2 and 3. Run conditions and results are given in Tables 4 and 5.

Contactor Optimization Tests

For each of the contactor designs tested, a statistically designed test matrix in the key operating parameters was employed. The experimental design was different for each contactor design; details are provided in a later section. At each test condition the pilot plant was operated until temperatures lined out. For a humidification efficiency test, data were collected for at least one hour after line out.

TEST SORBENT

The fresh sorbent used in all the recycle optimization tests was a high calcium hydrated lime obtained from Mississippi Lime Company. This lime was used in the 105 MW demonstration of the Coolside process and in previous tests of the Advanced Coolside process. A typical analysis of this lime is given in Table 1.

DISCUSSION OF RECYCLE OPTIMIZATION TEST RESULTS

PROCESS PERFORMANCE GOALS EXCEEDED

The results of the recycle optimization tests show that the process performance targets of 90% SO_2 removal and 60% sorbent utilization can be exceeded. Figure 3 shows the SO_2 removals and corresponding sorbent utilizations achieved in the recycle optimization tests at different combinations of process variables. The data show that the 90% SO_2 removal target can be achieved at sorbent utilizations of over 70%. The data also show that very high SO_2 removals (90 to 99+%) can be achieved while maintaining at least 60 % sorbent utilization. Sorbent recycle is a key to achieving these levels of performance.

RESULTS OF TESTS SIMULATING SO_2 REMOVAL WITH AN ESP

The tests listed in Tables 2 and 4 were conducted with hot air reheat to maintain a baghouse exit approach to saturation of ~25 °F, and with frequent baghouse pulse cleaning. This mode of operation was used to reduce the SO_2 removal in the baghouse and roughly simulate SO_2 removal with an ESP.

The results of these tests show that the target SO_2 removal of 90% can be achieved at sorbent utilizations of up to about 75%. In tests 12, 13 and 12A (Tables 2,4), conducted at various combinations of fresh Ca/S and recycle ratio, the system SO_2 removals were 90% for each test and the sorbent utilizations ranged from 60% to 75%. As shown in Table 2, the mode of operation limited the SO_2 removal in the baghouse to 3 to 7% absolute, which is within the approximate range that would be expected in an ESP. In-duct SO_2 removals for these tests ranged from 83 to 87%. The results of these tests indicate that with an ESP the system removal of 90% can be achieved.

RESULTS OF TESTS SIMULATING SO_2 REMOVAL WITH A BAGHOUSE

The tests listed in Tables 3 and 5 were conducted with a baghouse approach to saturation of 9-12 °F and less frequent baghouse pulsing (ca. every 30 to 60 min). These tests were conducted to simulate SO_2 removal in a plant with a baghouse. The results show that the process is capable of very high SO_2 removal (90 to greater than 99%), while maintaining the process target of 60% sorbent utilization. In tests 11 and 11A, at a 1.6 fresh Ca/S mol ratio and a recycle ratio of 3.8, the system SO_2 removals were 97 to 99+% and the sorbent utilizations were 60-61%. In test 17B at a fresh Ca/S ratio of 1.2 and a recycle ratio of 7 lb/lb, in-duct and system SO_2 removals were 84% and 92%, respectively.

EFFECT OF PROCESS VARIABLES

Moisture Addition to Recycle

The addition of moisture to the recycle sorbent had a strong positive effect on desulfurization performance of the sorbent. Figure 4 shows that the addition of 0.15 lb H_2O/lb of recycle sorbent, at a 1.2 fresh Ca/S mol ratio, a 5.0 recycle ratio and a 10 °F approach in the baghouse, increased the in-duct SO_2 removal from 59% to 81% and the system removal from 73% to 88% (Tests 6A and 10, Tables 3 and 5). The sorbent utilization increased from 61% with no moisture addition to 71% with moisture addition. Tests 6A, 7A and 8A, (Tables 3 and 5) also point out the positive effect of moisture addition. The system SO_2 removal (73%) and the sorbent utilization (61%) in Test 6A, with no moisture addition, were lower than the removals (81-84%) and utilizations (65-67%) in Tests 7A and 8A with moisture, even though Test 6A employed a higher recycle ratio.

Table 3 shows that there was little, if any, advantage in increasing the amount of moisture addition from 0.10 to 0.15 lb water/lb recycle sorbent. Test 8A made with 0.10 lb water and test 7A made with 0.15 lb water (other conditions the same) showed very similar SO_2 removals in the duct and system and very similar sorbent utilizations.

The optimum water addition level determined in pilot tests may not apply directly to large-scale operation. In the pilot plant the ratio of transport air to sorbent is much greater than typical for a large-scale transport system. Also, the air used in the pilot plant is dry plant air. Consequently, in the pilot plant more water is required on the sorbent to allow for the evaporation into the dry transport air.

Wetting/Injection of Recycle Sorbent Together With Fresh Lime

In the majority of the tests reported here, the fresh lime (dry) and wetted recycle sorbent were handled in separate feed systems. In some tests the fresh and recycle sorbents were blended, treated with water, and injected into the duct from one feeder. As shown in Tables 2 and 4, the wetting procedure had no apparent effect on the desulfurization performance of the sorbent. Test 12, conducted with separate injection, and test 12A, conducted with moisture addition to the combined sorbent feed, had essentially the same in-duct and system SO_2 removals and sorbent utilizations.

Fresh Ca/S Ratio and Recycle Ratio

Increasing the fresh Ca/S ratio or the recycle ratio increases the amount of available calcium in the system; that is, total calcium in the fresh and recycle sorbents not associated with sulfur ($Ca(OH)_2$, $CaCO_3$). An increase in total available calcium substantially increases the SO_2 removal, as shown in Figure 5. At a 10 °F approach to saturation in the baghouse and with 0.15 lb H_2O/lb recycle, increasing the total available Ca/S ratio from 2.3 to 3.8 increased the in-duct SO_2 removal from 60% to 88% and the system SO_2 removal from 84% to 97%. The data indicate that by maintaining a high enough concentration of available calcium in the sorbent, the process target of 90% SO_2 removal can be achieved or exceeded.

In-duct Residence Time

A study of in-duct residence time showed that there was little effect of residence time between 1.7 and 2.7 sec on the in-duct SO_2 removal. On the other hand, between 1.0 and 1.7 sec, residence time had a significant effect. The results indicate that high SO_2 removals and sorbent utilizations can be achieved with 1.7 to 2.0 sec in-duct residence time. Construction of additional ductwork to increase the residence time above 2.0 sec appears to be unwarranted.

At a 1.2 Ca/S mol ratio and a 7/1 recycle ratio (Test 13, Figure 6, Table 7), the in-duct removal increased from 62% at 1.0 s to 83% at 1.7-2.0 sec. At 2.7 sec the removal was 86%, a small increase over that observed at 1.7-2.0 sec. At a 1.5 Ca/S ratio and a 4.3 lb/lb recycle ratio (Test 12A, Figure 7, Table 7), there essentially was no effect of residence time in the range of 2.0 to 2.7 sec. The in-duct SO_2 removals were 83 and 84% at 2.0 and 2.7 sec, respectively.

The results presented in Tables 2, 3, 4 and 5 were obtained at an in-duct residence time of 2.7 sec. In two of the recycle simulation tests (12A and 13, Tables 2 and 4), residence times of 1.0, 1.7 and 2.0 sec also were studied to determine the effect on desulfurization performance of the sorbent. This was done by moving a gas sample probe to different locations in the duct between the sorbent injection location and the end of the duct. A stationary probe at the end of the duct measured results at 2.7 sec residence time. These measurements were made at steady-state recycle conditions.

PROCESS OPERABILITY OBSERVATIONS

Pilot plant operational experience during this test program was a positive indication for the operability and retrofit potential of the Advanced Coolside process. Although the pilot plant is not of sufficient scale to fully assess process operability, observations of pilot operation provide initial information on key operability issues. The pilot plant has been a useful tool in the past in identifying potential operability concerns. The recycle optimization tests discussed in this report involved over 15 months of pilot plant operation, including long-term tests of up to 115 hours in duration. Observations of different aspects of pilot plant operation are discussed below.

Operation With Wetted Recycle

Minimal operating problems were encountered in preparing, handling and feeding wetted recycle sorbent, as long as appropriate procedures and operating conditions were employed.

Duct Sorbent Injection at High Humidity

The pilot testing provided an opportunity to observe the effect of high humidity and duct configuration on operability. In all the pilot tests, flue gas at the contactor exit was at or near the saturation point (0 to 2 °F approach to saturation). As SO_2 capture proceeded, gas temperature and approach to saturation increased along the duct length. As shown in Figures 8 and 9, the ductwork between the sorbent injection point and the baghouse was comprised of 53 ft of 5-inch pipe, 24 ft of 8-inch pipe and 7 ft of 4-inch pipe. The gas velocity ranged from 18 to 58 ft/sec. There were seven locations where the flue gas changed direction, including one 180° bend. The residence time before the first 90° bend was less than 0.5 sec.

There were no major operating problems associated with high humidity flue gas conditions or with the many changes in flue gas flow direction. Operating procedures and conditions were selected to minimize deposition of wet solids in the ductwork. The conditions were selected based on previous CONSOL experience.

Overall, the operating results indicate that sorbent can be injected into very humid flue gas without significant operability problems. The results also show that it is possible to operate with changes in flue gas direction and with a short straight-run residence time after sorbent injection. This flexibility is an advantage for retrofit of the process.

Baghouse Operation

Baghouse operability was good at approach temperatures as low as 10 °F. There were no problems in removing the spent sorbent from bags or from the baghouse hopper. Baghouse operating procedures and conditions were selected based on prior CONSOL experience.

DATA RELIABILITY

The results presented here are for tests of relatively long duration for pilot plant optimization tests. As shown in Tables 4 and 5, test duration was generally over 20 hours, with one test lasting 115 hours. This duration assured that steady-state conditions were closely approached. The data reported represent data averaged over periods in which desulfurization performance and plant operation were lined out. The fact that performance was observed over an extended period of time increases the reliability of the performance data.

In each test, there was good agreement between sorbent utilization based on the continuous flue gas analyzer and the fresh and recycle sorbent feed/composition data and the utilization based on baghouse solids analyses. As shown in Tables 4 and 5, there was no more than 4% absolute difference between the two values in any test. The agreement between the two values confirms the accuracy of process flow and analyzer data. The methods for calculating utilization based on gas analyzer and solids data have been described in a earlier report.[1]

Tables 4 and 5 also show that there was good agreement between the utilization calculated assuming steady-state recycle conditions and that based on baghouse solids analysis. The steady-state value is simply the system SO_2 removal divided by the fresh Ca/S mol ratio. There was no more than 5% absolute difference between the two values in any test. The absolute average of the differences was 2%. This agreement indicates that steady-state recycle conditions were closely approached. It further confirms the accuracy of the process flow and analyzer data. This agreement is further illustrated in Figure 10.

To further confirm the desulfurization performance results, EPA Method 6 sampling tests were conducted on the flue gas during a recycle test (Test 11A, Table 5). This was done to compare the in-duct SO_2 removals measured by Method 6 with those measured by the continuous flue gas analysis system. Three tests were conducted with sampling at the contactor exit (prior to sorbent injection) and at the exit of the ductwork. As shown in Figure 11, there was generally good agreement between the two methods for measuring in-duct SO_2 removal. The only large discrepancy was in test 1, for which the EPA Method 6 sampling indicated an in-duct SO_2 removal of 96%, compared to 88% with the continuous gas analyzers. This difference may be attributed to the presence of spent sorbent observed in the Method 6 sample train filter during this test. In tests 2 and 3 the absolute difference in SO_2 removals was only 2-3%.

DISCUSSION OF CONTACTOR OPTIMIZATION

FIRST GENERATION CONTACTOR: WATERLOO SCRUBBER

Contactor Description

The contactor used in the initial pilot plant studies was a pilot Waterloo scrubber system (Figure 12) supplied by Turbotak, Inc., with a design throughput of 1000 acfm. It consisted of a preconditioning spray chamber, a modified centrifugal fan, and an entrainment separator (mist eliminator). Air atomizing nozzles were used to spray fine water droplets into the flue gas stream at the preconditioner inlet, at the preconditioner exit, and at the fan inlet.

The modified centrifugal fan was designed to further promote solid/liquid and gas/liquid contact. The centrifugal action of the fan forced slurry droplets to the fan housing for removal. An entrainment separator downstream of the fan removed any remaining droplets. Hydraulic nozzles in the entrainment separator prevented plugging by residual fly ash in the flue gas.

Humidification Efficiency

Fifty-two saturation efficiency tests (Table 9) were performed using the Waterloo scrubber. The first (HE-1) was run at the same spray nozzle water flows and air pressures as all previous tests in the pilot plant; these conditions were designed for submicron particle capture. The second (HE-2) was run at the same total water flow as the first and had about the same total droplet surface area as the first, but the water flows and air pressures to each nozzle were balanced. These two tests were considered baseline tests because the 1 gpm water flow and 30 μm Sauter mean droplet diameter were the same as used in all earlier pilot plant tests using the Waterloo scrubber. The results of the other 49 tests were compared to the first two tests. The low-air-pressure tests (HE-3 through HE-7) were all performed at the same 1 gpm water flow rate. Their pressures ranged from 15 to 35 psi. The total droplet surface area increased with increasing pressure. The tests with the largest droplets were conducted to explore the feasibility of using hydraulic nozzles; the largest droplet size tested was smaller than that which can be practically achieved with commercial hydraulic nozzles. Tests were conducted with the fan and fan nozzle shut off (Tests HE-8 through HE-12); the purpose of these tests was to explore the feasibility of eliminating the fan. These tests were performed using two different water flow rates: either the same gpm/nozzle as the previous tests (giving 0.66 gpm total) or the same total gpm as the previous tests (1 gpm total or 0.5 gpm/nozzle). The air pressures were varied to give approximately the same droplet surface area range as tests HE-1 through HE-7. The droplet sizes were estimated based on the air pressure and the water flow rate by using information supplied by the nozzle manufacturer.

Effect of Fan. The Waterloo scrubber fan was not required to achieve adequate humidification of the flue gas as long as sufficient water droplet surface area was maintained. The relative humidity at the contactor exit is shown in Figure 13 as a function of specific droplet surface area (m^2 droplet area/m^3 flue gas) for tests with and without the fan. The points must be compared on an equal surface area basis because two nozzles were used in the no-fan tests but three nozzles were used in the tests with the fan operating. Within experimental error, the relative humidity was the same with and without the fan at equivalent droplet surface area conditions. These results indicate that the fan is not necessary for the Advanced Coolside process; this simplifies the contactor design and significantly reduces capital cost.

Optimization of Operating Conditions. Results of the saturation efficiency tests indicate that the contactor operating conditions can be optimized. The high energy atomization used in the Waterloo scrubber for fine particulate control was not necessary to achieve near saturation conditions. Operation at lower atomization air pressures could significantly reduce contactor capital and operating costs.

The relative humidity was lower at lower atomizing air pressures when the water flow rate was held constant as shown by Figure 14. The triangles represent tests using three nozzles and the scrubber fan; the squares and diamonds represent two-nozzle, no-fan tests with the squares representing water flow rate of 0.33 gpm/nozzle and the diamonds 0.5 gpm/nozzle. The two triangles with the highest atomizing air pressure represent the design conditions for capture of submicron particles. The figure clearly shows that reduction of the air pressure reduced the contactor's ability to saturate the gas. The effect was small at the higher pressures but there appeared to be a critical pressure below which the effect was more significant. In tests using three nozzles, the humidity dropped off sharply when the pressure was less than about 25 psig. Using two nozzles, the humidity dropped off at around 45 psig.

This behavior is the result of droplet size and total droplet surface area. Larger droplets are produced if the atomizing air pressure is reduced while the water flow rate is held constant. As a result, there are comparatively fewer droplets to evaporate the same amount of water and less droplet surface area over which to do it. Figure 15 shows the same data as Figure 14 plotted as a function of mean drop diameter. Clearly, the smaller droplets gave better humidification.

Use of Hydraulic Nozzles. The feasibility of using hydraulic nozzles in place of the two-fluid nozzles was evaluated by examining the data from the tests producing the largest droplets. Commercial hydraulic nozzles generally produce droplets in the 80 to 1000 μm diameter range (Table 10) under practical operating conditions. In the Waterloo scrubber, droplets of ~60 μm Sauter mean diameter showed unacceptably low humidification efficiency (<90%). The smallest average droplet size produced by a commercial hydraulic nozzle is about 50 μm; however, the orifice size is so small that it would quickly plug unless the water were ultra-filtered. Also, the water flow rate is so low that the number of nozzles required for a commercial size installation would be impractical.

Fly Ash Collection Efficiency

Eight tests were conducted in which fly ash was injected into the gas stream ahead of the contactor to measure the particulate collection efficiency with different nozzle configurations. The scrubber fan was not in operation for these tests and the fan nozzle was not used. The variables were total water flow rate (0.4 to 0.9 gpm) and the atomizing air pressure (25 to 45 psig). Particulate removal efficiency was greater than 95% in all of the tests, indicating that the removal efficiency was not sensitive to the nozzle operating conditions over the ranges tested and that the scrubber fan was not needed to achieve particulate removal >90 wt %. The results are listed in Table 9.

Operability

Contactor operability was good throughout the tests. There were no problems with solids accumulation in the contactor. There were no problems with mist eliminator plugging using the recommended 1 gpm wash flow.

A short-term operability test was conducted using two spray nozzles with the fan off. Ash capture (>90 wt %) and humidification performance did not deteriorate during the test. The mist eliminator screens were washed periodically (about every half hour) to prevent the screens from plugging with the uncaptured fly ash.

SECOND GENERATION CONTACTOR: SIMPLIFIED TURBOTAK DESIGN

Description

CONSOL purchased from Turbotak a mechanically simpler second generation contactor for pilot plant testing (Figure 16). The final design was prepared by Turbotak Inc. based on CONSOL's recommendations. It consists of a redesigned contacting chamber and a mist eliminator; the fan was eliminated in the new design. The contact chamber employs four dual-fluid nozzles.

Humidification Efficiency

One hundred fifty tests were performed to verify the saturation efficiency of the second generation contactor and to identify the optimum nozzle operating conditions for economic flue gas saturation and fly ash removal (Table 11). The percent relative humidity was calculated (as described earlier in the "Test Procedures" section) based on the wet bulb temperature of the gas after dilution with hot air. This calculation sometimes gives a relative humidity greater than 100%; this is a result of the thermocouple inaccuracy: a 1 °F error in the thermocouple reading can lead to a ca. 5% error in the relative humidity. These values were interpreted as representing fully saturated (100% relative humidity) flue gas.

Many test conditions were identified that provided satisfactory saturation and ash removal but required less water and/or lower air pressures than the conditions recommended by Turbotak. These conditions would result in lower operating costs for the contactor. The humidification results for all 150 tests are plotted as relative humidity versus droplet surface area in Figure 17. The droplet surface area was varied by varying the water flows and pressures. The percent relative humidity was calculated, as described earlier, based on the wet bulb temperature at the baghouse after dilution with hot air. The droplet surface area was normalized relative to the surface area produced by the Turbotak design conditions. The vertical dotted line represents the surface area produced by the

Turbotak design; points to the left of this line were obtained in tests in which the droplet surface area was lower than the design. Clearly, a large number of the tests still provided sufficient humidification (>95% relative humidity, horizontal dotted line) at these operating conditions.

<u>Tests Using One, Two, or Three Spray Nozzles</u>. Sufficient humidification could be achieved with fewer than four nozzles operating at the design conditions. Various combinations of the nozzles were systematically evaluated. The nozzle that had the least effect on humidification performance was Nozzle 3. However, no optimization of the nozzle position or spray direction was attempted. Figure 18 shows the effect of operating fewer than four spray nozzles in the contactor. These data were obtained by operating the sprays at the design conditions with one, two, or three of the sprays turned off. The results of tests with all four sprays operating at the design conditions are shown for comparison.

The three-nozzle tests all gave good humidification performance (95 to 99% relative humidity). The poorest humidification was obtained in the one-nozzle tests. Nozzle 1 gave the best performance (63% relative humidity), followed by Nozzle 2 (49%) and Nozzle 4 (37%). A one-nozzle test was attempted with Nozzle 3, but the humidification performance poor and the test was aborted.

The humidification performance in the two-nozzle tests ranged from 61% relative humidity to 95% relative humidity at the contactor exit. The worst performance was achieved using Nozzles 3 and 4, the two worst performers in the one-nozzle tests. The best performance in the two-nozzle tests was obtained using Nozzles 1 and 2, the two best performers in the one-nozzle tests.

<u>Optimization of Contactor Operating Conditions</u>. The effect of reducing atomizing air and water flow is shown in Figure 19. The relative humidity of the flue gas exiting the contactor is plotted as a function of the total water flow along lines of constant atomizing air pressure. A reduction in air pressure requires a reduction in the water flow rate to keep the relative humidity sufficiently high, or the droplets become too large to provide adequate surface area for evaporation. The figure shows that the water flow rate and the atomizing air pressure can be reduced from the design conditions without significantly reducing

the saturation efficiency. An optimum nozzle operating condition of 30 psig air pressure to each nozzle and 0.6 gpm/1000 acfm total water flow was chosen based on these results. The optimized operating conditions gave similar humidification performance and fly ash removals as the original design operation conditions. At ~500 scfm (730 acfm) flue gas flow, both original and optimized operating conditions allowed close to 100% relative humidity; the fly ash capture averaged 97% for the original design conditions and 93% for the optimized design conditions. At ca. 700 scfm (1025 acfm) the relative humidity averaged 96% for the original operating conditions and 94% for the optimized conditions; the fly ash capture averaged 83% for the original conditions and 85% for the optimized conditions. The optimized operating conditions will reduce capital and operating costs, as a result of the reduced air pressure (lower compressor capital cost and operating energy) and the reduced water flow (less pumping and wastewater handling requirements). This optimized contactor operating condition was used in the subsequent recycle tests and in the sorbent optimization tests.

The contactor was designed by Turbotak to minimize ash deposition at the wet/dry interface. Tests were conducted to evaluate operability. These tests showed that the operating procedures and conditions recommended by Turbotak minimized ash deposition with minimum extra water usage.

<u>Effect of Contactor Gas Throughput</u>. The humidification was somewhat lower at higher contactor flue gas throughputs, over a range of 700 to 1500 acfm. This is a result of the lower residence time of the flue gas in the contactor and less droplet surface area per volume of flue gas for evaporation. Figure 20 shows the trend for tests in which the nozzle air pressure was 30 psig or more using all four nozzles. Optimization tests were not performed for high contactor gas throughput; this program was canceled to allow testing of the third generation contactor.

<u>Fly Ash Collection Efficiency</u>

Fourteen tests were conducted at ca. 500 scfm (730 acfm) in which fly ash was injected into the gas stream ahead of the contactor to measure the particulate collection efficiency using different nozzle configurations. The variables were total water flow rate (0.4 to 1.6 gpm) and the atomizing air pressure (15 to 50 psig). Particulate removal efficiency was greater than 90% in all of the tests, indicating that the removal efficiency was not sensitive to the nozzle operating conditions over the ranges tested. The results are listed in Table 12.

At ca. 700 scfm (1025 acfm), two tests were conducted. The fly ash capture efficiency dropped to 83 to 85% at the higher flow rate. This is still sufficient fly ash capture for the process, since the contactor is designed to be installed ahead of an existing particulate collector.

Operability

Contactor operability was good throughout the tests. There were no problems with solids accumulation in the contactor, nor were there problems with mist eliminator plugging. At shutdown, the mist eliminator was clean, with no indication of any solids build-up.

During a long-term (115 on-stream hours) test of sorbent recycle, the simplified contactor operated without difficulty.

Downstream Desulfurization Performance

The majority of the tests in the recycle sorbent optimization program discussed above were conducted with the second generation contactor at the conditions identified as optimum in the contactor optimization tests. The desulfurization performance targets were exceeded. Some tests were conducted with the original Waterloo scrubber. Both contactors were operated to achieve near saturation conditions and no difference in downstream desulfurization performance was observed.

THIRD GENERATION CONTACTOR DESIGN: IN-DUCT VENTURI + CYCLONE

Description

The third generation contactor (Figure 21) was designed for lower capital cost and a reduced plant footprint. It consists of a low-pressure-drop, in-duct venturi followed by cyclonic separator. Water is sprayed by hydraulic nozzles at the throat of the venturi. The venturi reduces water droplet size and provides turbulent contact between droplets and flue gas for efficient particle capture and humidification. The water/fly ash mix is separated from the flue gas by the downstream separator.

The design pressure drop for the venturi and separator is 5" WC. The design water requirement is about 5 gal/1000 acf. These are higher than for the second generation contactor design (~1.5" WC and 1 gal/1000 scf); however, the third generation design is significantly smaller and has a significantly lower capital cost. Furthermore, the use of hydraulic nozzles instead of two-fluid nozzles can save capital and operating costs for air compression. A detailed cost analysis is presented in Topical Report No. 6, Conceptual Commercial Design and Economic Evaluation.

The venturi contactor installed in the pilot plant was purchased by CONSOL from Fisher-Klosterman, Inc. Because of the small scale of the venturi, there was not a gradual expansion section after the throat for pressure recovery; this increased the permanent pressure loss compared to a large scale unit.

Initial testing of the venturi contactor as supplied by the vendor indicated that there was difficulty in achieving acceptable humidification efficiency, because of the very short contact time between the venturi throat and the cyclone in the small-scale unit. Some tests were conducted with water spraying upstream of the venturi to increase the residence time; this improved humidification somewhat. Eventually, the contactor was modified for increased contact time downstream of the venturi throat. The contact time downstream of the throat is critical for humidification, because the water droplet size is reduced in the throat. The modified design better simulates a full-scale unit, because in a larger unit the transition between the venturi throat and the cyclonic separator would be longer and provide more residence time. To modify the pilot plant contactor, a 6 ft section was added between the venturi throat and the cyclonic separator. This increased the residence time between the throat and the separator to about 0.1 sec at full load.

Humidification Efficiency

The humidification testing indicated that the initial design of the venturi contactor gave unacceptable humidification performance. Modification of this design to increase the contact time downstream of the venturi throat allowed reasonably close approaches (~1 to 4 °F) to be achieved. Injection of steam at the exit of the cyclonic separator allowed near saturation conditions to be achieved at all flue gas flows. A small amount of fine mist carry-over from the separator also incrementally lowered the approach. Exploratory tests indicated that near saturation can be achieved by using two-fluid nozzles upstream of the separator instead of a venturi to generate small droplets. Detailed results of the these humidification tests are discussed below.

Original Venturi Contactor Design. Table 13 summarizes humidification efficiency tests with the venturi contactor as received from Fisher-Klosterman. With this design, acceptable humidification for the Advanced Coolside process was not achieved.

In a test at the design conditions, ca. 1150 acfm flue gas flow, 5 gpm water at the venturi throat and 5" WC pressure drop across the contactor, the approach to saturation was about 25 °F. In tests at lower flue gas flows (700 to 770 acfm), the approach was about 14 °F. The closer approach to saturation with a lower flue gas flow suggests that humidification was limited by the liquid/gas contact time.

As a simple means of increasing the liquid/gas contact time in the contactor, hydraulic nozzles were added about 10 ft upstream of the venturi throat. This increased the contact time to about 0.2 s at full load. The test data in Table 14 show that this improved humidification significantly. However, a close approach to saturation was only achieved in a few tests with low flue gas flow rates, high liquid/gas ratios and high venturi pressure drops. In tests at flue gas flows ranging from ~300 to 500 acfm and water flows of around 5 gpm split between the throat and upstream nozzles, the approach to saturation ranged about 1 to 5 °F. In tests at higher flows, the approach to saturation varied over a range of about 5 to 15 °F, depending on process conditions (Table 14). Again, the positive effect of reduced gas flow on the humidity suggests that humidification performance was limited by the liquid/gas contact time.
Although a close approach was achieved in some of the tests with upstream water spraying, this design was not considered to be optimum. The ability to operate the contactor at higher load was desired to minimize contactor size. Furthermore, the above design did not take full advantage of the significant reduction in water droplet size which occurs across the venturi.

Modified Venturi Contactor. As discussed above, the venturi contactor was modified by CONSOL for improved humidification performance. This involved addition of a 6 ft duct section between the venturi throat and the cyclonic separator. The initial tests discussed above indicated that performance was limited by liquid/gas contact time. This modification increased the flue gas residence time between the venturi throat and the cyclonic separator to ~0.1 sec at full load. This design makes more effective use of the venturi than the use of upstream nozzles, because it increases the contact time after the reduction in water droplet size by the venturi. The modified design showed improved humidification efficiency compared to the original design. Reasonably close approach to saturation was achieved.

Table 15 summarizes the humidification tests conducted with the modified venturi contactor. The modified contactor was capable of achieving approaches to saturation in the range of 1 to 4 °F at flue gas flow rates of ranging from about 650 to 920 acfm. As shown in the table, numerous combinations of operating conditions were tested. In some tests the water spraying was split between the venturi throat and supplemental nozzles either upstream of the throat or between the throat and the separator. As shown, different combinations of conditions were identified that achieved approaches to saturation in the 1 to 4 °F range. In general, the approach tended to be closer at lower flue gas flows, 650 to 750 acfm, and higher at flow rates around 900 acfm. The use of supplemental nozzles upstream or downstream of the venturi throat had a small positive effect allowing reasonably close approaches to be achieved at the higher flow rates.

Tests of Mist Carry-Over from the Contactor. A series of tests was conducted simulating increased levels of fine mist carry-over from the modified venturi contactor (Table 16). These tests were conducted by adding small amounts (0.025 to 0.075 gpm) of mist generated by a high pressure (>80 psig) two-fluid atomizer at the exit of the cyclonic separator. These data may be important because a larger scale separator will not be as efficient as the small-scale unit in capturing the finest droplets. Also, the use of a small two-fluid nozzle may be feasible to lower the approach closer to the saturation point. As shown in the table, the approach to saturation in the downstream duct could be lowered to near 0 °F.

Modified Contactor with Steam Injection. Injection of low quality steam at the exit of the separator in the modified venturi contactor was explored as a means of incrementally lowering approach to near the saturation point. The data in Table 17 show that with steam injection the approach can be lowered to 0 to 2 °F with a full load flue gas flow of ca. 1000 acfm. This range of approaches indicates saturation within the range of uncertainty of the measurements. The tests in Table 17 were conducted with water spraying at the venturi throat only. Desulfurization tests discussed below indicate that steam injection incrementally improved desulfurization performance, a further indication of a closer approach to saturation. Another possible effect of steam injection is condensation on the surface of the injected sorbent. The steam injection rates used ranged from 0.05 to 0.5 lb/min. As shown in Table 17, the optimum steam injection rate may be in the middle of this range; however, the optimum is difficult to determine because of the experimental uncertainty in measuring very close approaches.

Tests of an Alternative Contactor Configuration. A few exploratory tests were conducted to evaluate an alternative contactor configuration. This configuration involved in-duct water spraying with no venturi, followed by a cyclonic separator. A hydraulic nozzle was used at the duct inlet to quench the flue gas and reduce problems of a wet/dry interface. Two-fluid nozzles were used downstream to generate fine droplets and achieve a close approach. This design avoids the gas pressure drop associated with the venturi but requires air compressors for atomization. As shown in Table 18, the brief testing of this concept indicated that approaches to saturation in the 0 to 2 °F could be achieved.

Fly Ash Collection Efficiency

Fly ash collection efficiency was consistently over 99% in four pilot plant tests conducted with EPA Method 17 sampling. This collection efficiency exceeds the target of 90% desired to reduce fly ash in the sorbent recycle loop. Table 19 gives contactor operating conditions and performance data for these tests.

Fly ash collection efficiency was independent of gas flow over a range of 380 to 1025 acfm. The results indicate that a single venturi contactor can handle the range of turndown required for a commercial application to follow changing boiler load. This is an important result, since the first conceptual commercial design assumed that two parallel contactors would be required to handle load changes.

Operability

Operability of the third generation contactor was good throughout the performance tests. There were no problems with fly ash accumulation in the venturi, on the spray nozzles, in the cyclonic separator or in the ductwork. In the initial tests where water was sprayed upstream of the venturi throat, there was a small amount of solids dropout at the bottom of the horizontal inlet duct; however, this accumulation leveled off after a short period of operation. With the modified design there was little dropout in the horizontal ductwork between the venturi and the separator.

Desulfurization Results

Recycle tests with sorbent injection downstream of the third generation contactor (venturi contactor) indicated that performance targets could be met with sorbent addition levels 5 to 15% higher than with the second generation contactor. With the use of steam injection at the separator exit, with the use of supplemental nozzles or with slightly increased mist carry-over from the cyclonic separator, about 5% more sorbent was used to achieve 90% removal. This difference approaches the range of uncertainty in the pilot plant measurements. In earlier tests where the approach to saturation ranged 1 to 4 °F, 10 to 15% more sorbent was required, presumably due to the slightly lower humidity. All of the desulfurization tests were conducted with the modified design in which the contact time after the venturi throat was increased to 0.1 sec. Based on the test results, the use of steam injection at the separator appears to be the preferred mode of operation.

Once-Through Tests. Initial once-through (no recycle) tests showed that SO_2 removals with the venturi contactor were similar to those obtained with the second generation contactor (Turbotak spray chamber and mist eliminator). Figure 22 shows that the in-duct SO_2 removal at a Ca/S mol ratio of 1.5 was 51% with both the venturi contactor and the second generation Turbotak system. At a 2.0 Ca/S ratio the in-duct removals with the venturi contactor and the Turbotak system were 58 and 62%, respectively.

Initial Recycle Tests. Initial recycle simulation tests with the modified venturi contactor showed somewhat less efficient desulfurization than obtained with the first or second generation contactors. This is likely a result of the somewhat higher approach to saturation with the venturi contactor, 2-4 °F, compared to that with previous designs (<1 °F).

Detailed test conditions and results for the recycle tests are given in Table 20, and the summarized results are given in Table 21. Test 23, made at a 1.38 fresh Ca/S mol ratio and a recycle ratio of 6.2 (dry), gave in-duct and system SO_2 removals of 83 and 92%, respectively. With the second generation contactor, this level of SO_2 removal was obtained with less fresh sorbent. Test 13 (Table 21) gave in-duct and system SO_2 removals of 87 and 90% at a 1.21 fresh Ca/S ratio and a recycle ratio of 7.0. The approach to saturation in the duct downstream of the venturi (Test 23) was 2 to 4 °F, compared to near saturation with the second generation contactor (Test 13).

Three different approaches were employed to increase humidification efficiency and improve desulfurization with the venturi contactor. The first approach involved the use of supplemental nozzles located upstream of the venturi throat, in addition to the hydraulic nozzles at the venturi throat. The second involved the use of two-fluid nozzle(s) at the cyclone exit to inject very small amounts of fine mist; this simulated increased mist carry-over. The third involved addition of small amounts of steam downstream of the cyclonic separator, with the venturi throat nozzles in operation. In addition, a test was conducted to explore possible enhancement by additives. These approaches are discussed below.

Tests with Supplemental Nozzles. In an attempt to supply additional flue gas/water contact to improve flue gas saturation, tests were conducted with two-fluid nozzles installed 5-10 ft upstream of the venturi throat. These additional nozzles lowered the approach to saturation somewhat; however, they had little, if any, effect on desulfurization performance. Tests 25-1 and 25-2 (Table 21), which were made at a 1.32 fresh Ca/S mol ratio, gave in-duct and system SO_2 removals of 81 and 85%, respectively. In Test 25-3, conducted at a 1.40 fresh Ca/S ratio, in-duct and system SO_2 removals were 85 and 92%, essentially the same as in Test 23 at a 1.38 fresh Ca/S ratio with no additional nozzles.

Tests with Increased Mist at the Contactor Exit. These tests were conducted primarily to explore the effect of slightly increased mist carry-over from the cyclonic separator. The small-scale cyclonic separator was likely more efficient in removing the finest water droplets than a larger scale unit. The separator may also have been more effective than the mist eliminator used in the second generation contactor design; this could partly account for the somewhat better desulfurization performance obtained with the second-generation design.

Employing a high-pressure (90 psig) two-fluid atomization nozzle to produce a fine mist at the exit of the cyclonic separator showed significant positive results. As discussed previously, the approach to saturation was lowered to near saturation. Tests 25-5 and 25-6, conducted with a 1.33 fresh Ca/S ratio gave in-duct SO_2 removals as high as 86%, slightly better than the 83% removal at the higher Ca/S ratio of 1.38 without mist injection (Test 23). The system SO_2 removal in tests 25-5 and 25-6 was 89%. The amount of mist injected was small, 0.0025 to 0.0090 gpm. In tests at higher mist injection levels there was an operability problem at the sorbent injection location. Sorbent was injected at a 90° bend downstream of the separator; apparently some droplets and sorbent impacted the duct wall at the bend.

The above tests indicate a positive effect of a small amount of fine mist downstream of the contactor. A small supplemental nozzle at the separator exit may be a means of incrementally improving desulfurization performance, if it can be adequately controlled to prevent operability problems.

Test with Additive Addition. Test 23B was conducted at a 1.30 fresh Ca/S mol ratio with a small concentration of NaCl added to the recycle sorbent (0.02 Na/Ca mol). No additional nozzles were used. The additive had a small positive effect on baghouse SO_2 removal, and essentially no effect on in-duct removal. The in-duct and system SO_2 removals in test 23B were 76 and 89%, respectively. A more detailed discussion on the effects of additives is given in Topical Report 3.

Tests with Steam Addition at the Separator Exit. As reported previously, injection of a small amount of low-quality steam at the exit of the cyclonic separator lowered the approach to saturation from a range of 1 to 4 °F to near saturation. Desulfurization tests indicated that steam injection also improved desulfurization performance.

Detailed test conditions and results for the recycle tests with steam addition are given in Table 22, and the summarized results are given in Table 23. Test 27, conducted at a 1.26 fresh Ca/S mol ratio, a recycle ratio of 7, 0.12 lb water/lb recycle, and 20 lb/hr steam addition, gave in-duct and system SO_2 removals of 80 and 91%, respectively. These removals are comparable to those obtained with the second generation contactor at these conditions, i.e., test 17B (Table 3) gave SO_2 removals of 84 and 92%.

Test 26 (Table 23) was made with steam addition and no water added to the recycle sorbent. The objective was to determine whether use of steam injection could eliminate the need for the water addition step. A 90% SO_2 removal efficiency was obtained in the system; however, a higher fresh Ca/S mol ratio (1.41) was required. The in-duct SO_2 removal was 77%. The duct exit approach to saturation was 5-7 °F higher in test 26 than seen in the tests where moisture was added to the recycle sorbent; this would account for the poorer desulfurization performance.

Test 28 was made with steam addition and with a small concentration (0.004 wt %) of hydrochloric acid (HCl) in the recycle treatment water. This simulated commercial operation where the contactor recycle water would be used to treat the recycle sorbent. This water would pick up chloride ion from the coal combustion flue gas. This mode of operation is preferred because it tightens the process water balance and reduces the chloride concentration in the contactor recirculation loop. The HCl showed a small enhancing effect on the SO_2 removal efficiency in the duct, and had no effect in the baghouse. The in-duct and system SO_2 removals in test 28 were 86 and 91%, compared to 80 and 91% in test 27 with no HCl addition. This was different from the effect of NaCl, as reported in Topical Report No. 3.

Tests LT-01A, B, and C were made with steam addition and a lower approach to saturation in the baghouse (4-6 °F). The tests demonstrated the high SO_2 removal efficiencies possible using a baghouse as the solids collection device. At a

1.26 fresh Ca/S mol ratio and a 6.7 recycle ratio, the system SO_2 removals ranged from 92 to 97%. The in-duct removals for these tests were 67 to 82%.

OPTIMIZATION OF RECYCLE SORBENT MOISTURE ADDITION EQUIPMENT

A test was conducted in which the recycle sorbent was wetted using a pilot-scale, continuous pugmill. Performance of the pugmill was compared to that of the high-intensity mixer (Littleford) used in previous pilot plant tests. The results from this test indicated that a pugmill can produce a satisfactory product both from a materials handling standpoint and from a reactivity standpoint. These results are encouraging because a pugmill has substantially lower capital and operating cost than high intensity mixer.

The pugmill test was conducted at Heyl & Patterson's (H&P) Pittsburgh, Pennsylvania, test facility. A 600 lb batch of recycle material was prepared by combining spent sorbent from several pilot plant runs. Half of the 600 lb batch of recycle material was combined in the pugmill with 0.16 lb of fresh Mississippi lime per lb of recycle and 0.12 lb of water per lb of recycle material. The remaining 300 lb of recycle material was combined with fresh hydrate and water in the proportions indicated using the high intensity Littleford mixer.

The pugmill was a continuous, pilot-scale (100-1200 lb/hr) unit. Two feeders were available - a small screw feeder and a larger capacity vibrating feeder. The small unit was used to feed the fresh, hydrated lime. The larger unit proved to be unreliable for feeding the recycle material. This material became aerated with handling and bridged in the feed hopper. As a result, the recycle material was fed by hand.

Initial tests were conducted with fresh hydrated lime only. Observations during these tests led to the installation of a second spray nozzle for more uniform water addition. These nozzles were positioned in the middle of the mill with the first immediately after the discharge of the screw feeder and the second approximately half way down the length of the pugmill. This arrangement yielded a product that was visibly uniform with 10-15% of the material in -½ inch lumps that broke readily upon handling.

After a satisfactory product was produced with the fresh hydrate, the pugmill was cleaned thoroughly and the screw feeder purged of hydrate. Feeding of the recycle solids was initiated simultaneously with feeding of the Mississippi lime.

After a layer of dry solids was established part way down the pugmill bed, water addition was initiated. The entire 300 lb of recycle material was fed to the pugmill, combined with 48 lb of fresh hydrate and water, and the product collected in 3 drums within 25 min. Average solid residence time in the pugmill was estimated at 30 sec. The product produced had the same appearance as described above and about 10% water according to the moisture determination conducted at H&P.

This product was evaluated in lab tests and compared with the similar recycle blend prepared in the Littleford, high-intensity mixer. Lab analyses of grab samples from different periods of pugmill operation (Table 24) confirmed that the material produced was homogeneous with respect to moisture content and composition (e.g., calcium and sulfur contents).

The product was evaluated in tests in the Advanced Coolside pilot plant and compared with the similar recycle blend prepared in the Littleford, high-intensity mixer. The results obtained from testing the pugmill-produced material are very encouraging. No operability problems were encountered either in the pilot plant feed system or in the reaction duct. The desulfurization performance was identical, within experimental error, to that obtained with the material produced by the Littleford mixer, as shown in Table 25. These pilot plant tests were not carried out to steady-state recycle conditions. The tests were merely run long enough (2 hr) to compare the reactivity of the two feedstocks.

These results indicate that a pugmill can be used in the Advanced Coolside process, resulting in capital and operating cost savings in the recycle sorbent pre-treatment step.

OTHER DESIGN OPTIMIZATION WORK

In addition to pilot plant optimization testing discussed above, engineering studies were conducted to explore process improvements in all major process subsystems, including the sorbent handling, recycle handling, flue gas handling and waste handling systems. These engineering studies are discussed in detail in Topical Report No. 6, Conceptual Commercial Design and Economic Evaluation. Key areas identified for process improvement/cost reduction include:

- Use of hydrocyclones instead of a thickener to concentrate the fly ash slurry before mixing with spent sorbent.
- Use of on-site lime hydration of quicklime for larger plants.
- Simplification of the flue gas reheat system.
- Improvements in the recycle handling system design.
- Simplification of the ductwork conceptual design.

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TABLE 1

TYPICAL ANALYSIS OF MISSISSIPPI HYDRATED LIME

BET Surface Area, m^2g	24.4
Lime Index	93.9
As-Received Lime, wt %	
Moisture	0.5
Ash (925 °C)	75.1
Carbonate (CO_3)	2.2
$Ca(OH)_2$ (TGA)	90.9
Ash Elementals, As-Received Lime, wt %	
SiO_2	0.8
Al_2O_3	0.1
TiO_2	<0.1
Fe_2O_3	0.1
CaO	74.7
MgO	0.6
Na_2O	<0.1
K_2O	<0.1
P_2O_5	<0.1
SO_3	0.2
Malvern Particle Size, wt %	
+ 66.9 μm	0
66.9 x 42.9	1.1
42.9 20.5	6.3
20.5 x 11.4	9.3
11.4 x 5.4	31.3
5.4 x 1.9	40.9
-1.9	11.1
Mean Particle dia., μm	5.3

TABLE 2

SUMMARY OF RECYCLE TEST RESULTS: BAGHOUSE APPROACH OF 23-24 °F

Test	Fresh Ca/S, mol	Recycle Ratio (a)	lb Water per lb Recycle Sorbent	Total $Ca(OH)_2/S$ mol Ratio	Baghouse Approach, °F	SO_2 Removal, %		Sorbent Util., %	
						Duct	System (b)	Steady State (d)	Solids Analyses
12	1.4	4.5	0.15	2.2	23	83	90	63	62
13	1.2	6.9	0.12	2.1	23	87	90	75	70
12A (c)	1.5	4.3	0.15	2.5	24	84	90	60	59

Common Conditions: SO_2 Inlet Concentration = 1500 ppm (dry); Flue Gas Flow = 340 SCFM

(a) lb dry recycle/lb fresh lime
(b) duct + baghouse
(c) fresh lime and recycle sorbent wetted together and fed by one feeder
(d) calculated steady-state sorbent utilization

TABLE 3

SUMMARY OF RECYCLE TEST RESULTS: BAGHOUSE APPROACH OF 9-12 °F

Test	Fresh Ca/S, mol	Recycle Ratio (a)	lb Water per lb Recycle Sorbent	Total $Ca(OH)_2/S$ mol Ratio	Baghouse Approach, °F	SO_2 Removal, %		Sorbent Util., %	
						Duct	System (b)	Steady State (d)	Solids Analyses
6A	1.2	5.0	0.00	2.2	10	59	73	61	58
7A	1.3	3.3	0.15	1.8	9	60	84	67	68
8A	1.2	3.4	0.10	1.8	11	64	81	65	66
9	1.5	3.5	0.15	2.2	12	70	90	61	63
10	1.2	4.9	0.15	1.7	9	81	88	71	68
11	1.6	3.9	0.15	2.4	11	91	97	60	58
11A	1.6	3.8	0.15	2.4	12	88	100	61	61
17B (c)	1.2	6.9	0.12	1.4	10	84	92	76	72

Common Conditions: SO_2 Inlet Concentration = 1500 ppm (dry)

(a) lb dry recycle/lb fresh lime
(b) duct + baghouse
(c) fresh lime and recycle sorbent wetted together and fed by one feeder
(d) calculated steady-state sorbent utilization

TABLE 4

TEST CONDITIONS AND RESULTS, RECYCLE SIMULATION TESTS, BAGHOUSE APPROACH OF 23-24 °F

Test	12	13	12A (a)
Run Time, hr	34	115	12
Sorbent Data			
Fresh Ca/S Mole Ratio (b)	1.43	1.21	1.49
Fresh Feedrate, lb/hr (c)	7.45	6.29	7.77
Recycle Feedrate, lb/hr	41.67	52.92	41.39
Recycle Ratio, lb recycle/lb fresh lime	5.59	8.41	5.33
Recycle Ratio, dry basis	4.46	6.89	4.30
Recycle Available Ca/S, mol ratio (b)	1.93	1.71	2.05
Total Available Ca/S, mol ratio (b)	3.36	2.92	3.54
Water Addition, lb/hr	5.79	5.89	5.76
lb Water/lb Recycle Sorbent	0.15	0.12	0.15
Duct Flue Gas Conditions			
In–Duct Residence Time, s	2.7	2.7	2.7
Duct Inlet SO_2 Content, ppmv–dry	1501	1501	1500
<u>Approach to Saturation, °F</u>			
Duct Exit	5	4	5
Baghouse Exit	23	23	24
Solids Loading, gr/scf	16.9	20.3	16.9
Contactor Inlet Temp, °F	281	281	280
Contactor Exit Temp, °F	130	130	128
Duct Exit Temp, °F	132	131	132
Baghouse Exit Temp, °F	149	149	150
Baghouse Exit Wet Bulb, °F	126	126	126
Duct Inlet Flue Gas Flow, scfm	340	340	340
SO_2 Removal, %			
In–Duct	83	87	84
System (Duct + Baghouse)	90	90	90
Sorbent Utilization, %			
Steady State (d)	63	75	60
Flue Gas Analyzers (e)	59	69	57
Ash Analysis (f)	63	69	59
TGA Analysis (g)	61	71	58

(a) Fresh and recycle sorbents treated and fed together.
(b) Includes all calcium in fresh and recycle feeds not associated with sulfur (e.g., $Ca(OH)_2$ and $CaCO_3$).
(c) Fresh feed was Mississippi hydrated lime.
(d) Based on flue gas analysis and fresh sorbent feed rate/composition, assuming steady state.
(e) Based on flue gas analysis and recycle and fresh sorbent feed rates/compositions.
(f) Based on baghouse solids analysis (S, Ca).
(g) Based on baghouse solids analysis (TGA).

TABLE 5

TEST CONDITIONS AND RESULTS, RECYCLE SIMULATION TESTS, BAGHOUSE APPROACH OF 9-12 °F

Test	6A	7A	8A	9	10	11	11A	17B (a)
Run Time, hr	26	29	36	26	22	23	23	64
Sorbent Data								
Fresh Ca/S Mole Ratio (b)	1.20	1.25	1.23	1.47	1.23	1.62	1.62	1.22
Fresh Feedrate, lb/hr (c)	6.20	6.40	6.40	7.86	6.40	8.40	8.44	6.37
Recycle Feedrate, lb/hr	33.49	27.06	25.86	34.36	39.89	41.75	40.93	53.54
Recycle Ratio, lb recycle/lb fresh lime	5.40	4.23	4.04	4.37	6.23	4.97	4.85	8.41
Recycle Ratio, dry basis	4.96	3.33	3.38	3.47	4.94	3.94	3.82	6.89
Recycle Available Ca/S, mol ratio (b)	1.85	1.10	1.05	1.54	1.40	2.17	1.86	1.40
Total Available Ca/S, mol ratio (b)	3.05	2.35	2.29	3.01	2.63	3.79	3.48	2.62
Water Addition, lb/hr	0.00	3.53	2.35	4.54	5.33	5.80	5.66	5.96
lb Water/lb Recycle Sorbent	0.00	0.15	0.10	0.15	0.15	0.15	0.15	0.12
Duct Flue Gas Conditions								
In–Duct Residence Time, s	2.7	2.7	2.7	2.7	2.7	2.7	2.7	2.7
Duct Inlet SO_2 Content, ppmv–dry	1489	1477	1495	1542	1504	1499	1501	1500
Approach to Saturation, °F								
Duct Exit	8	6	7	7	5	6	6	2
Baghouse Exit	10	9	11	12	9	11	12	10
Solids Loading, gr/scf	13.6	11.5	11.1	14.5	15.9	17.2	16.9	20.6
Contactor Inlet Temp, °F	299	300	299	299	300	299	282	281
Contactor Exit Temp, °F	131	130	130	129	129	131	130	129
Duct Exit Temp, °F	135	133	133	132	131	132	132	130
Baghouse Exit Temp, °F	136	135	136	136	134	136	137	137
Baghouse Exit Wet Bulb, °F	126	126	125	125	125	125	125	127
Duct Inlet Flue Gas Flow, scfm	340	340	340	340	340	340	340	340
SO_2 Removal, %								
In–Duct	59	60	64	70	81	91	88	84
System (Duct + Baghouse)	73	84	81	90	88	97	100	92
Sorbent Utilization, %								
Steady State (d)	61	67	65	61	71	60	61	76
Flue Gas Analyzers (e)	58	64	64	59	67	56	60	73
Ash Analysis (f)	58	68	66	63	68	58	61	72
TGA Analysis (g)	57	68	61	65	65	62	61	73

(a) Fresh and recycle sorbents treated and fed together.
(b) Includes all calcium in fresh and recycle feeds not associated with sulfur (e.g., $Ca(OH)_2$ and $CaCO_3$).
(c) Fresh feed was Mississippi hydrated lime.
(d) Based on flue gas analysis and fresh sorbent feed rate/composition, assuming steady state.
(e) Based on flue gas analysis and recycle and fresh sorbent feed rates/compositions.
(f) Based on baghouse solids analysis (S, Ca).
(g) Based on baghouse solids analysis (TGA).

TABLE 7

EFFECT OF RESIDENCE TIME ON SO_2 REMOVAL,
RECYCLE SIMULATION TESTS

Common Conditions:

SO_2 Inlet Concentration = 1500 ppm (dry)
Sorbent = Mississippi hydrated lime

Test	12A	13
Fresh Ca/S, mol	1.5	1.2
Recycle Ratio (a)	4.3	6.9
lb Water/lb Recycle	0.15	0.12
Duct Approach, °F	5	4
Baghouse Approach, °F	24	23
SO_2 Removal, %		
Duct Residence Time, s		
1.0	-	62
1.7	-	83
2.0	83	82
2.7	84	86 (avg)
System	90	90

(a) lb dry recycle/lb fresh lime

TABLE 9

HUMIDIFICATION EFFICIENCY TESTS USING FIRST GENERATION CONTACTOR (WATERLOO SCRUBBER)

Run DOE–HE–	1	2	3	4	5	6	7	8	9	10	11	12	13
Rx Flow (std m^3/s)	0.146	0.145	0.145	0.144	0.145	0.145	0.145	0.145	0.145	0.146	0.145	0.145	0.145
WL Flow (std m^3/s)	0.247	0.247	0.247	0.247	0.247	0.248	0.247	0.247	0.246	0.246	0.246	0.244	0.239
React Inlet Press., "H2O	−8.95	−9.21	−9.46	−9.76	−9.58	−9.51	−9.52	−10.48	−10.84	−11.15	−11.39	−11.40	−9.19
WL Exit Pressure, "H2O	−7.15	−7.26	−7.53	−7.77	−7.58	−7.43	−7.43	−10.25	−10.49	−10.39	−10.44	−10.33	−7.20
TEMPERATURES, °F													
Flue Gas	430	430	421	410	420	420	420	415	413	414	415	419	411
WL Flue Gas	281	281	282	280	282	281	282	283	281	281	279	281	282
WL Inlet	280	280	281	279	280	280	280	282	280	280	280	280	280
WL Blower In	128	128	130	132	130	129	129	133	131	131	130	128	128
WL Exit	127	127	129	131	128	128	127	132	128	129	128	127	125
WL Exit 2	127	127	129	131	128	128	128	132	130	130	128	127	129
WL Exit Aft Bypass	128	127	129	131	128	128	128	132	132	131	129	128	129
Rx Before Sol	127	127	129	130	128	128	128	132	131	130	129	128	128
Rx Inlet	127	127	129	131	128	128	128	132	132	131	129	128	129
Avg Rx Skin Temp	127	127	128	130	129	129	128	130	131	131	129	128	131
Rx Exit	128	128	130	131	129	129	129	131	132	132	130	129	130
BH Inlet	140	139	140	141	139	139	141	140	140	140	138	137	137
Recycle FG	148	148	152	153	150	150	150	154	152	153	152	151	141
BH Exit	150	150	151	152	149	149	149	151	152	151	150	149	146
BH Exit Wet Bulb	125	125	125	125	125	125	125	125	125	126	125	125	126
Sup Heat Air Temp	577	578	577	577	538	558	558	558	558	558	558	559	753
WL Fan ON/OFF	ON	ON	ON	ON	ON	ON	ON	OFF	OFF	OFF	OFF	OFF	ON
WL Exit O2	12.6	12.3	12.6	12.7	12.3	12.5	12.4	12.3	12.4	12.5	12.4	12.4	11.1
BH Exit O2	13.1	12.8	13.1	13.1	12.8	13.0	12.9	12.7	12.9	12.9	12.9	12.8	11.4
Nozzle Flow, gpm													
1	0.40	0.33	0.33	0.33	0.33	0.33	0.33	0.33	0.33	0.50	0.50	0.33	0.38
2	0.40	0.33	0.33	0.33	0.33	0.33	0.33	0.33	0.33	0.50	0.50	0.33	0.38
3	0.20	0.33	0.33	0.33	0.33	0.33	0.33	0.00	0.00	0.00	0.00	0.00	0.38
Total gal/min	1.00	0.99	0.99	0.99	0.99	0.99	0.99	0.66	0.66	1.00	1.00	0.66	1.13
Nozzle Pressure, psig													
1	40	45	20	15	25	30	35	35	45	45	60	60	35
2	40	45	20	15	25	30	35	35	45	45	60	60	35
3	60	45	20	15	25	30	35						35
# of nozzles used	3	3	3	3	3	3	3	2	2	2	2	2	3
Total Atom. Air, kg/sec	0.00833	0.00645	0.00273	0.00222	0.00371	0.00427	0.00508	0.00323	0.00421	0.00345	0.00489	0.00628	0.00544
kg air / kg water	0.1322	0.1034	0.0438	0.0355	0.0595	0.0684	0.0814	0.0777	0.1011	0.0547	0.0775	0.1510	0.0767
RELATIVE HUMIDITY, %													
WL Inlet	2.46	2.46	2.36	2.46	2.46	2.48	2.46	2.38	2.44	2.50	2.49	2.42	2.52
WL Exit	95.2	95.2	88.8	84.1	92.9	93.3	94.1	82.1	91.5	89.9	92.6	94.6	103.2
BH Exit	49.3	49.5	47.6	46.0	51.0	50.4	50.7	47.9	46.8	47.9	49.2	49.9	56.0
Fly Ash Collection, wt%													

TABLE 9 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING FIRST GENERATION CONTACTOR (WATERLOO SCRUBBER)

Run DOE–HE–	14	15	16	17	18	19	20	21	22	23	24	25	26
Rx Flow (std m³/s)	0.144	0.145	0.144	0.144	0.144	0.144	0.144	0.143	0.143	0.144	0.144	0.144	0.144
WL Flow (std m³/s)	0.239	0.239	0.198	0.200	0.203	0.208	0.213	0.249	0.249	0.248	0.251	0.251	0.248
React Inlet Press., "H2O	−9.51	−8.54	−9.21	−9.58	−9.11	−9.54	−9.70	−8.35	−8.57	−6.67	−6.80	−6.64	−6.83
WL Exit Pressure, "H2O	−7.12	−6.63	−7.32	−7.66	−7.21	−7.62	−7.84	−6.47	−6.66	−4.80	−4.92	−4.72	−4.92
TEMPERATURES, °F													
Flue Gas	416	427	416	412	425	426	421	425	419	416	416	421	408
WL Flue Gas	281	282	281	282	281	279	281	281	281	281	282	281	281
WL Inlet	280	281	280	280	279	277	279	280	279	280	281	280	280
WL Blower In	128	128	129	130	128	129	130	129	127	132	132	129	137
WL Exit	125	124	126	127	126	127	128	127	125	131	131	127	139
WL Exit 2	129	128	128	130	128	128	130	130	128	131	131	129	137
WL Exit Aft Bypass	129	128	128	129	128	129	129	130	130	131	131	130	136
Rx Before Sol	128	127	128	127	127	127	126	129	128	133	133	133	133
Rx Inlet	129	128	129	129	129	129	129	131	131	132	132	131	135
Avg Rx Skin Temp	131	131	131	132	131	131	131	132	132	132	132	132	132
Rx Exit	130	130	131	131	131	130	131	132	132	133	133	132	135
BH Inlet	137	137	139	139	139	138	139	139	139	140	140	140	142
Recycle FG	141	140	142	143	141	142	143	142	141	146	145	142	151
BH Exit	147	146	150	150	150	150	150	150	149	151	151	151	152
BH Exit Wet Bulb	126	125	125	125	125	125	125	126	124	126	125	125	125
Sup Heat Air Temp	775	776	621	618	618	618	618	555	555	618	618	619	618
WL Fan ON/OFF	ON	ON	ON	ON	ON	ON	ON	ON	ON	OFF	OFF	OFF	OFF
WL Exit O2	11.7	12.7	13.2	12.4	13.5	12.9	12.6	12.2	11.5	12.0	11.6	11.8	11.3
BH Exit O2	11.8	12.7	13.6	12.8	13.8	13.3	13.0	12.5	11.9	12.4	12.1	12.3	11.8
Nozzle Flow, gpm													
1	0.25	0.20	0.45	0.45	0.30	0.20	0.30	0.25	0.38	0.25	0.38	0.30	0.30
2	0.25	0.20	0.45	0.45	0.30	0.20	0.30	0.25	0.38	0.25	0.38	0.30	0.30
3	0.25	0.20	0.45	0.45	0.30	0.20	0.30	0.25	0.38	0.00	0.00	0.00	
Total gal/min	0.75	0.60	1.35	1.35	0.90	0.60	0.90	0.75	1.13	0.50	0.75	0.60	0.60
Nozzle Pressure, psig													
1	35	45	45	25	45	25	25	35	35	35	35	45	25
2	35	45	45	25	45	25	25	35	35	35	35	45	25
3	35	45	45	25	45	25	25	35	35	0	0	0	
# of nozzles used	3	3	3	3	3	3	3	3	3	2	2	2	2
Total Atom. Air, kg/sec	0.00693	0.00960	0.00602	0.00320	0.00790	0.00549	0.00432	0.00667	0.00519	0.00447	0.00346	0.00517	0.00291
kg air / kg water	0.1466	0.2537	0.0708	0.0376	0.1392	0.1451	0.0762	0.1411	0.0732	0.1418	0.0731	0.1368	0.0769
RELATIVE HUMIDITY, %													
WL Inlet	2.56	2.41	2.40	2.45	2.42	2.63	2.51	2.53	2.39	2.56	2.39	2.40	2.48
WL Exit	103.6	104.1	97.4	95.1	97.4	96.2	92.5	96.3	98.3	85.6	83.4	94.4	65.9
BH Exit	55.3	54.5	48.8	49.3	48.6	49.8	49.9	51.0	49.5	49.0	47.4	47.6	46.4
Fly Ash Collection, wt%													

TABLE 9 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING FIRST GENERATION CONTACTOR (WATERLOO SCRUBBER)

Run DOE–HE–	27	28	29	29.1	30	31	32	33	34	35	36	37	38
Rx Flow (std m³/s)	0.144	0.144	0.144	0.144	0.144	0.144	0.144	0.144	0.144	0.144	0.144	0.144	0.144
WL Flow (std m³/s)	0.251	0.250	0.251	0.250	0.250	0.250	0.250	0.250	0.250	0.250	0.250	0.250	0.232
React Inlet Press., "H2O	−6.53	−6.83	−6.86	−6.84	−7.06	−5.44	−5.57	−5.45	−5.57	−5.50	−5.44	−5.57	−5.63
WL Exit Pressure, "H2O	−4.67	−4.89	−4.95	−4.95	−5.12	−3.61	−3.60	−3.55	−3.59	−3.55	−3.57	−3.55	−3.66
TEMPERATURES, °F													
Flue Gas	423	406	409	415	403	412	416	417	415	406	402	411	413
WL Flue Gas	281	281	280	281	281	282	282	283	283	284	282	283	283
WL Inlet	280	280	279	280	281	280	280	280	280	281	280	281	280
WL Blower In	129	138	133	131	137	130	130	129	130	139	138	135	132
WL Exit	127	140	127	128	138	130	130	128	128	141	139	133	132
WL Exit 2	129	137	129	129	135	131	131	129	130	138	137	134	132
WL Exit Aft Bypass	130	136	131	130	134	131	131	130	130	137	137	138	136
Rx Before Sol	131	133	132	133	133	131	132	132	132	135	137	136	136
Rx Inlet	131	135	132	131	134	132	131	131	132	135	136	138	135
Avg Rx Skin Temp	132	132	132	132	132	132	132	132	132	132	135	138	134
Rx Exit	132	135	132	132	134	132	132	132	133	135	136	139	136
BH Inlet	140	142	140	140	141	139	139	139	140	142	143	145	141
Recycle FG	142	151	145	143	150	144	144	142	143	152	151	148	145
BH Exit	151	153	151	151	152	150	150	150	150	152	153	155	150
BH Exit Wet Bulb	125	125	125	125	124	125	126	126	126	126	126	126	127
Sup Heat Air Temp	618	618	619	619	618	589	589	589	589	589	589	589	509
WL Fan ON/OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF
WL Exit O2	11.8	11.5	11.5	11.4	11.0	12.2	12.4	12.5	12.4	12.0	12.1	12.4	12.7
BH Exit O2	12.3	12.0	11.6	11.9	11.6	12.7	12.9	12.9	12.9	12.5	12.5	12.8	13.0
Nozzle Flow, gpm													
1	0.20	0.20	0.45	0.45	0.45	0.38	0.25	0.30	0.45	0.30	0.45	0.30	0.30
2	0.20	0.20	0.45	0.45	0.45	0.38	0.25	0.45	0.30	0.45	0.30	0.30	0.30
3													
Total gal/min	0.40	0.40	0.90	0.90	0.90	0.75	0.50	0.75	0.75	0.75	0.75	0.60	0.60
Nozzle Pressure, psig													
1	45	25	45	45	25	35	35	45	45	25	25	45	25
2	45	25	45	45	25	35	35	45	45	25	25	25	45
3													
# of nozzles used	2	2	2	2	2	2	2	2	2	2	2	2	2
Total Atom. Air, kg/sec	0.00650	0.00363	0.00390	0.00390	0.00216	0.00342	0.00429	0.00505	0.00450	0.00238	0.00247	0.00391	0.00397
kg air / kg water	0.2577	0.1438	0.0688	0.0687	0.0381	0.0723	0.1360	0.1068	0.0952	0.0503	0.0522	0.1034	0.1051
RELATIVE HUMIDITY, %													
WL Inlet	2.37	2.41	2.41	2.45	2.33	2.45	2.50	2.57	2.58	2.55	2.61	2.57	2.65
WL Exit	94.3	64.8	93.4	92.9	67.4	87.1	87.5	96.2	95.5	66.0	69.5	83.0	86.3
BH Exit	47.4	45.4	46.8	48.2	45.6	48.9	49.9	51.1	50.8	48.6	47.7	44.6	52.2
Fly Ash Collection, wt%	97.2	96.75	95.8		95.4	97.05		96.6	96.8				

TABLE 9 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING FIRST GENERATION CONTACTOR (WATERLOO SCRUBBER)

Run DOE–HE–	39	40	41	42	43	44	45	46	47	48	49	50	51
Rx Flow (std m^3/s)	0.144	0.144	0.143	0.143	0.143	0.143	0.143	0.143	0.143	0.143	0.144	0.142	0.142
WL Flow (std m^3/s)	0.227	0.227	0.250	0.250	0.250	0.248	0.250	0.249	0.249	0.248	0.250	0.248	0.248
React Inlet Press., "H2O	–5.57	–5.65	–5.67	–5.70	–5.54	–6.14	–5.57	–6.21	–6.26	–5.89	–5.98	–6.50	–6.51
WL Exit Pressure, "H2O	–3.52	–3.64	–3.63	–3.65	–3.55	–4.18	–3.52	–4.26	–4.33	–3.81	–3.91	–4.73	–4.75
TEMPERATURES, °F													
Flue Gas	405	406	406	412	401	412	403	413	412	403	403	428	429
WL Flue Gas	283	283	283	284	283	283	284	281	280	288	286	285	285
WL Inlet	280	280	280	281	279	280	281	279	278	283	281	281	281
WL Blower In	133	132	133	131	133	133	134	131	129	140	136	130	133
WL Exit	133	131	132	131	133	132	134	129	127	141	135	130	134
WL Exit 2	133	132	132	131	132	132	134	130	129	139	134	130	132
WL Exit Aft Bypass	137	136	132	132	133	132	134	131	131	140	138	132	133
Rx Before Sol	134	134	133	133	134	131	135	131	131	138	137	131	132
Rx Inlet	135	134	132	131	133	131	134	131	131	139	137	132	132
Avg Rx Skin Temp	134	135	132	132	132	132	133	132	133	135	135	133	133
Rx Exit	136	135	133	133	134	133	134	132	133	139	138	133	133
BH Inlet	142	141	140	140	141	140	141	139	140	145	144	140	141
Recycle FG	146	145	147	146	147	146	148	144	143	153	149	143	146
BH Exit	150	150	150	150	151	150	151	150	150	153	152	149	149
BH Exit Wet Bulb	127	126	126	126	127	127	126	126	126	126	125	126	126
Sup Heat Air Temp	509	509	589	589	589	589	588	588	589	588	589	589	589
WL Fan ON/OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF	OFF
WL Exit O2	12.6	12.3	12.7	12.8	12.3	12.9	12.3	12.7	12.6	13.2	13.1	13.1	13.4
BH Exit O2	12.9	12.7	13.0	13.2	12.7	13.1	12.7	13.2	13.2	13.4	13.4	13.4	13.7
Nozzle Flow, gpm													
1	0.45	0.45	0.35	0.20	0.50	0.20	0.50	0.20	0.50	0.20	0.50	0.35	0.20
2	0.45	0.45	0.35	0.20	0.50	0.20	0.50	0.50	0.20	0.50	0.20	0.35	0.50
3													
Total gal/min	0.90	0.90	0.70	0.40	1.00	0.40	1.00	0.70	0.70	0.70	0.70	0.70	0.70
Nozzle Pressure, psig													
1	45	25	35	25	25	45	45	45	45	25	25	35	45
2	25	45	35	45	45	25	25	45	45	25	25	35	25
3													
# of nozzles used	2	2	2	2	2	2	2	2	2	2	2	2	2
Total Atom. Air, kg/sec	0.00312	0.00302	0.00345	0.00496	0.00278	0.00475	0.00285	0.00486	0.00673	0.00274	0.00274	0.00409	0.00460
kg air / kg water	0.0549	0.0531	0.0782	0.1967	0.0440	0.1885	0.0452	0.1102	0.1524	0.0622	0.0621	0.0926	0.1043
RELATIVE HUMIDITY, %													
WL Inlet	2.62	2.50	2.61	2.50	2.68	2.64	2.56	2.65	2.62	2.43	2.34	2.52	2.54
WL Exit	82.0	85.3	85.1	86.0	83.9	85.3	81.0	91.8	96.9	64.9	73.6	89.2	80.8
BH Exit	51.6	50.2	51.1	50.5	50.6	51.8	50.2	51.4	49.4	47.4	45.8	51.3	52.0
Fly Ash Collection, wt%													97.2

TABLE 11

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-01	NC-01.1	NC-02	NC-02.1	NC-03	NC-04	NC-05	NC-06	NC-07	NC-08	NC-09	NC-10	NC-11	NC-12	NC-13	NC-14
Nozzle Pressure, psig																
1	40	40	40	40	35	35	45	35	25	20	15	40	45	35	25	20
2	50	50	50	50	35	35	45	35	25	20	15	50	45	35	25	20
3	50	50	50	50			45	35	25	20	15	50	45	35	25	20
4	45	45	45	45	35		45	35	25	20	15	45	45	35	25	20
Nozzle Flow, gpm																
1	0.43	0.43	0.43	0.43	0.38	0.35	0.20	0.20	0.20	0.20	0.20	0.43	0.40	0.40	0.40	0.40
2	0.20	0.20	0.20	0.20	0.38	0.35	0.20	0.20	0.20	0.20	0.20	0.20	0.40	0.40	0.40	0.40
3	0.20	0.20	0.20	0.20			0.20	0.20	0.20	0.20	0.20	0.20	0.40	0.40	0.40	0.40
4	0.30	0.30	0.30	0.30	0.38		0.20	0.20	0.20	0.20	0.20	0.30	0.40	0.40	0.40	0.40
Total gal/min	1.13	1.13	1.13	1.13	1.13	0.70	0.80	0.80	0.80	0.80	0.80	1.13	1.60	1.60	1.60	1.60
No. of nozzles used	4	4	4	4	3	2	4	4	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	2	0	0	0	0	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	308	305	308	308	305	305	305	308	306	305	305	305	300	301	300	300
Flue Gas °F	436	437	436	422	426	424	422	419	411	407	404	407	419	412	413	412
Rx Inlet °F	132	132	132	132	132	133	132	132	132	132	133	132	132	132	132	132
Rx Exit °F	133	133	133	133	133	133	133	133	133	133	134	133	133	133	133	133
BH Inlet °F	137	136	137	136	136	137	137	137	137	137	137	137	137	137	136	136
BH Exit °F	150	150	150	150	150	150	150	150	150	150	150	150	150	150	150	150
BH Exit Wet Bulb °F	125	124	125	126	125	125	125	125	127	125	126	126	126	127	126	125
React Inlet Press., "H2O	−5.10	−5.05	−5.00	−4.70	−5.40	−5.54	−4.97	−5.01	−5.13	−5.26	−5.29	−5.04	−4.71	−4.67	−4.85	−5.00
WL Exit O2 vol %	14.0	13.9	14.2	13.4	12.1	12.1	14.5	14.2	13.1	12.9	12.7	13.9	14.0	13.5	12.7	12.4
BH Exit O2 vol %	14.5	14.5	14.7	14.0	12.9	13.0	14.6	14.3	13.5	13.3	13.1	14.3	16.6	16.7	16.7	16.7
WL Flow (std ft³/min)	518	512	518	513	528	528	480	483	482	485	483	480	479	476	479	484
Room Temp. °F	71	74	76	72	73	76	71	70	71	75	78	80	71	69	71	76
WL Inlet °F	280	281	280	280	280	281	281	281	282	281	282	279	281	281	280	281
WL Exit °F	126	124	125	127	127	129	125	125	128	128	132	126	125	127	128	132
WL Exit 2 °F	129	128	128	129	129	130	128	128	130	130	131	129	129	130	130	132
WL Exit Aft Bypass °F	132	131	132	132	132	132	132	131	132	132	133	132	132	131	131	132
Recycle FG °F	140	140	140	142	141	143	141	141	143	143	146	143	142	143	143	145
Sup Heat Air Temp °F	305	306	306	305	306	305	305	305	305	305	306	306	305	305	305	305
WL Exit Pressure, "H2O	−3.12	−3.09	−3.04	−2.69	−3.42	−3.50	−2.89	−3.12	−3.10	−3.24	−3.24	−3.00	−2.87	−2.88	−2.99	−3.10
Rel Hum WL Exit	82.5	80.7	82.1	84.6	80.6	79.8	81.5	81.4	84.6	81.2	83.0	82.7	81.8	86.4	84.6	80.3
Rel Hum BH Exit	50.1	49.1	49.7	52.1	50.2	50.0	48.3	48.4	52.0	50.2	50.4	51.2	48.7	50.4	49.5	48.3
Avg Rx Skin Temp °F	132.4	132.6	132.7	132.2	132.4	132.6	132.8	132.6	132.6	132.4	132.9	132.5	132.5	132.4	132.1	132.0
Barometric Pressure "Hg	28.85	28.85	28.85	28.85	28.85	28.85	28.70	28.70	28.70	28.70	28.70	28.70	28.85	28.85	28.85	28.85
BH Exit Conditions:																
Dry Bulb Temp.	150	150	150	150	150	150	150	150	150	150	150	150	150	150	150	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	1.959	1.905	1.946	2.010	1.941	1.935	1.957	1.950	2.029	1.957	2.006	2.000	1.971	2.024	1.967	1.950
p(t) =	3.693	3.669	3.683	3.682	3.666	3.668	3.722	3.713	3.693	3.687	3.707	3.692	3.747	3.739	3.706	3.713
H =	0.08570	0.08258	0.08494	0.08874	0.08471	0.08436	0.08550	0.08509	0.08982	0.08557	0.08856	0.08812	0.08628	0.08938	0.08614	0.08512
press corr for H =	0.004228	0.004072	0.004186	0.004364	0.004180	0.004167	0.004824	0.004801	0.005062	0.004828	0.004906	0.004906	0.004253	0.004402	0.004244	0.004199
p =	1.846	1.787	1.832	1.903	1.827	1.821	1.843	1.836	1.924	1.845	1.900	1.892	1.856	1.914	1.854	1.835
% RH =	50.0	48.7	49.7	51.7	49.8	49.6	49.5	49.4	52.1	50.0	51.3	51.3	49.5	51.2	50.0	49.4
% of gas from SupHtr	7.94	7.81	7.86	7.92	7.78	7.72	8.01	7.97	7.88	7.89	7.81	7.88	8.06	8.22	8.26	8.06
WL Exit Conditions:																
H =	0.09769	0.09400	0.09672	0.10112	0.09639	0.09593	0.09819	0.09767	0.10299	0.09815	0.10149	0.10105	0.09647	0.10218	0.09852	0.09717
p =	1.983	1.918	1.966	2.043	1.960	1.952	1.981	1.972	2.064	1.981	2.038	2.031	1.997	2.061	1.997	1.974
p(t) =	1.984	1.900	1.948	2.036	2.014	2.142	1.935	1.929	2.074	2.083	2.313	1.993	1.955	2.028	2.079	2.328
% RH =	99.9	101.0	100.9	100.3	97.3	91.1	102.4	102.3	99.5	95.1	88.1	101.9	102.1	101.6	96.1	84.8

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-15	NC-16	NC-17	NC-18	NC-19	NC-20	NC-21	NC-22
Nozzle Pressure, psig								
1	15	40	35	35	35	35	35	35
2	15	50	35	35	35	35	35	35
3	15	50	35	35	35	35	35	35
4	15	45	35	35	35	35	35	35
Nozzle Flow, gpm								
1	0.40	0.43	0.40	0.40	0.40	0.40	0.40	0.40
2	0.40	0.20	0.20	0.20	0.20	0.20	0.40	0.40
3	0.40	0.20	0.20	0.20	0.40	0.40	0.20	0.20
4	0.40	0.30	0.20	0.40	0.20	0.40	0.20	0.40
Total gal/min	1.60	1.13	1.00	1.20	1.20	1.40	1.20	1.40
No. of nozzles used	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0
Rx Flow (scfm)	301	304	304	305	305	304	304	305
Flue Gas °F	400	421	421	418	417	419	418	430
Rx Inlet °F	133	133	132	133	132	132	132	132
Rx Exit °F	134	134	134	134	133	133	133	133
BH Inlet °F	137	137	137	137	136	136	136	136
BH Exit °F	151	150	150	150	149	149	149	150
BH Exit Wet Bulb °F	125	126	126	126	126	126	126	126
React Inlet Press., "H2O	−5.01	−4.76	−4.79	−4.75	−4.80	−4.80	−4.92	−4.74
WL Exit O2 vol %	12.5	14.3	14.5	14.0	13.8	13.6	13.6	13.6
BH Exit O2 vol %	16.8	17.0	17.0	17.1	17.0	17.1	17.1	17.8
WL Flow (std ft³/min)	476	477	476	475	477	482	480	495
Room Temp. °F	82	78	75	75	75	78	75	80
WL Inlet °F	281	280	281	281	279	280	280	281
WL Exit °F	134	126	126	127	126	126	126	126
WL Exit 2 °F	134	130	130	130	128	128	128	128
WL Exit Aft Bypass °F	134	132	132	132	131	131	131	131
Recycle FG °F	149	144	144	144	142	141	141	142
Sup Heat Air Temp °F	306	306	305	306	305	305	305	306
WL Exit Pressure, "H2O	−3.10	−2.87	−2.88	−2.85	−2.89	−2.91	−2.96	−2.90
Rel Hum WL Exit	79.5	81.6	81.8	83.2	83.7	84.0	84.2	85.2
Rel Hum BH Exit	47.8	50.7	50.9	52.0	51.4	51.3	51.7	49.4
Avg Rx Skin Temp °F	132.9	133.0	133.1	133.0	132.1	132.0	132.3	132.3
Barometric Pressure "Hg	28.85	28.70	28.70	28.70	28.70	28.70	28.70	28.70
BH Exit Conditions:								
Dry Bulb Temp.	151	150	150	150	149	149	149	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	1.942	1.977	1.977	2.010	1.979	1.974	1.966	1.990
p(l) =	3.751	3.689	3.680	3.673	3.629	3.623	3.619	3.718
H =	0.08452	0.08679	0.08682	0.08876	0.08704	0.08679	0.08752	0.08744
press corr for H =	0.004175	0.004886	0.004886	0.004982	0.004891	0.004875	0.004918	0.004926
p =	1.824	1.867	1.868	1.904	1.872	1.867	1.881	1.880
% RH =	48.6	50.6	50.8	51.8	51.6	51.5	52.0	50.6
% of gas from SupHtr	7.76	7.50	7.52	7.53	7.62	7.63	7.58	8.11
WL Exit Conditions:								
H =	0.09616	0.09911	0.09916	0.10139	0.09952	0.09923	0.10001	0.10053
p =	1.956	1.997	1.998	2.037	2.004	1.999	2.013	2.022
p(l) =	2.487	1.975	1.974	2.027	1.995	1.993	2.001	2.011
% RH =	78.7	101.1	101.3	100.5	100.5	100.3	100.6	100.5

Run No.	NC-23	NC-24	NC-25	NC-26	NC-27	NC-28	NC-29	NC-30
Nozzle Pressure, psig								
1	40	35	35	35	35	35	35	40
2	50	35	35	35	35	35	35	50
3	50	35	35	35	35	35	35	50
4	45	35	35	35	35	35	35	45
Nozzle Flow, gpm								
1	0.43	0.20	0.20	0.20	0.20	0.20	0.20	0.43
2	0.20	0.20	0.20	0.20	0.20	0.40	0.40	0.20
3	0.20	0.20	0.20	0.40	0.40	0.20	0.20	0.20
4	0.30	0.20	0.40	0.20	0.40	0.20	0.40	0.30
Total gal/min	1.13	0.80	1.00	1.00	1.20	1.00	1.20	1.13
No. of nozzles used	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0
Wash Water, gpm	2	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0
Rx Flow (scfm)	305	305	306	305	300	300	300	299
Flue Gas °F	428	418	418	416	428	428	426	428
Rx Inlet °F	132	129	128	128	129	129	130	129
Rx Exit °F	133	131	130	130	131	131	131	131
BH Inlet °F	136	135	134	133	134	134	134	134
BH Exit °F	150	150	149	149	150	150	150	150
BH Exit Wet Bulb °F	126	126	125	125	125	126	127	126
React Inlet Press., "H2O	−4.50	−4.63	−4.71	−4.77	−4.85	−4.79	−4.73	−4.82
WL Exit O2 vol %	13.8	14.1	14.0	14.0	14.0	14.0	13.9	13.8
BH Exit O2 vol %	17.8	15.9	15.8	15.9	16.4	16.5	16.7	16.6
WL Flow (std ft³/min)	499	472	473	473	495	489	497	496
Room Temp. °F	79	79	82	84	75	75	73	74
WL Inlet °F	281	280	280	280	281	281	279	280
WL Exit °F	126	126	124	125	125	126	128	126
WL Exit 2 °F	128	128	126	127	127	128	128	128
WL Exit Aft Bypass °F	131	129	127	128	129	129	129	129
Recycle FG °F	142	142	141	142	141	142	142	142
Sup Heat Air Temp °F	306	306	306	306	305	305	305	305
WL Exit Pressure, "H2O	−2.55	−2.76	−2.83	−2.87	−2.96	−2.88	−2.85	−2.94
Rel Hum WL Exit	83.9	91.3	90.3	91.9	88.2	91.4	94.2	91.7
Rel Hum BH Exit	49.3	51.9	49.5	50.2	49.4	50.2	51.4	50.8
Avg Rx Skin Temp °F	132.1	131.8	131.0	130.7	130.5	130.4	130.9	131.2
Barometric Pressure "Hg	28.70	28.70	28.70	28.70	28.78	28.78	28.78	28.78
BH Exit Conditions:								
Dry Bulb Temp.	150	150	149	149	150	150	150	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	1.984	1.993	1.917	1.938	1.933	1.996	2.046	2.001
p(l) =	3.713	3.656	3.632	3.643	3.680	3.712	3.713	3.696
H =	0.08710	0.08782	0.08343	0.08463	0.08423	0.08782	0.09080	0.08815
press corr for H =	0.004901	0.004934	0.004691	0.004759	0.004426	0.004814	0.004768	0.004630
p =	1.873	1.886	1.804	1.827	1.819	1.886	1.941	1.892
% RH =	50.5	51.6	49.7	50.2	49.4	50.8	52.3	51.2
% of gas from SupHtr	8.12	8.74	9.02	9.23	9.10	9.34	9.18	9.22
WL Exit Conditions:								
H =	0.10013	0.10163	0.09686	0.09648	0.09752	0.10195	0.10523	0.10219
p =	2.015	2.041	1.958	1.966	1.975	2.052	2.108	2.056
p(l) =	1.960	1.994	1.891	1.915	1.920	1.991	2.073	2.009
% RH =	101.8	102.4	103.6	103.7	102.9	103.1	101.7	102.3

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-03.1	NC-04.1	NC-31	NC-32	NC-33	NC-34	NC-35	NC-36	NC-37	NC-38	NC-39	NC-40	NC-41	NC-42	NC-43	NC-44
Nozzle Pressure, psig																
1	35	35	35	40	25	25	30	30	35	25	25	25	25	25	25	25
2	35	35	35	50	40	40	45	45	35	25	25	25	25	45	45	45
3			35	50	40	40	45	45	35	25	25	45	45	25	25	45
4	35		35	45	25	25	30	30	35	25	45	25	45	25	45	25
Nozzle Flow, gpm																
1	0.38	0.35	0.20	0.20	0.20	0.20	0.30	0.30	0.25	0.40	0.40	0.40	0.40	0.40	0.40	0.40
2	0.38	0.35	0.10	0.10	0.10	0.10	0.15	0.15	0.25	0.40	0.40	0.40	0.40	0.10	0.10	0.10
3			0.10	0.10	0.10	0.10	0.15	0.15	0.25	0.40	0.40	0.10	0.10	0.40	0.40	0.10
4	0.38		0.15	0.15	0.15	0.15	0.23	0.23	0.25	0.40	0.10	0.40	0.10	0.40	0.10	0.40
Total gal/min	1.13	0.70	0.55	0.55	0.55	0.55	0.83	0.83	1.00	1.60	1.30	1.30	1.00	1.30	1.00	1.00
No. of nozzles used	3	2	4	4	4	4	4	4	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	1	0	1	0	2	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	299	299	305	304	304	304	304	304	304	309	309	309	309	310	309	309
Flue Gas °F	421	419	423	427	422	426	426	428	425	404	408	407	414	408	410	412
Rx Inlet °F	130	130	131	130	130	131	131	131	131	130	130	130	130	130	130	130
Rx Exit °F	131	131	131	131	131	132	131	132	132	131	131	131	131	131	131	131
BH Inlet °F	134	135	135	135	135	135	135	135	135	134	134	134	134	134	134	134
BH Exit °F	150	150	150	150	150	150	149	150	149	149	149	149	149	149	149	149
BH Exit Wet Bulb °F	127	126	126	126	126	125	125	125	126	126	125	125	125	125	125	125
React Inlet Press., "H2O	−5.09	−5.17	−4.88	−4.58	−5.03	−5.09	−5.10	−5.03	−5.22	−5.08	−5.11	−5.08	−5.12	−5.15	−5.20	−5.15
WL Exit O2 vol %	12.6	12.6	14.1	14.7	13.8	13.7	13.7	13.6	12.9	12.6	12.8	13.2	13.2	13.0	13.2	13.1
BH Exit O2 vol %	16.7	16.7	17.2	17.2	17.3	17.2	17.2	17.2	13.6	12.9	13.2	13.5	13.6	13.4	13.6	13.4
WL Flow (std ft³/min)	503	498	485	488	488	505	500	519	510	487	486	488	488	487	488	488
Room Temp. °F	76	79	72	71	72	72	76	78	80	73	71	71	72	77	80	82
WL Inlet °F	280	280	280	281	279	280	280	280	280	281	280	280	280	281	280	280
WL Exit °F	127	128	127	127	127	128	128	125	127	128	125	128	128	128	128	128
WL Exit 2 °F	129	129	129	129	129	128	129	128	129	130	128	129	129	130	128	128
WL Exit Aft Bypass °F	129	130	130	130	130	130	130	130	130	130	129	129	129	130	130	129
Recycle FG °F	143	144	143	143	143	143	143	142	143	145	143	143	143	144	143	143
Sup Heat Air Temp °F	305	305	305	305	305	305	305	305	305	306	306	305	305	306	306	306
WL Exit Pressure, "H2O	−3.16	−3.25	−2.93	−2.70	−3.06	−3.09	−3.15	−3.08	−3.26	−3.08	−3.13	−3.10	−3.12	−3.17	−3.19	−3.12
Rel Hum WL Exit	92.0	87.4	87.5	86.6	87.7	83.8	84.5	83.9	88.2	90.1	87.7	88.7	88.5	87.8	87.4	87.6
Rel Hum BH Exit	51.2	49.6	51.6	51.8	51.4	49.9	50.0	49.7	51.5	51.2	49.7	50.4	50.3	50.7	50.4	50.2
Avg Rx Skin Temp °F	130.9	131.1	131.2	131.1	131.1	131.3	131.2	131.3	131.3	131.3	131.5	130.8	131.0	130.4	131.0	131.1
Barometric Pressure "Hg	28.78	28.78	28.75	28.75	28.75	28.75	28.75	28.75	28.75	28.60	28.60	28.60	28.60	28.60	28.60	28.60
BH Exit Conditions:																
Dry Bulb Temp.	150	150	150	150	150	150	149	150	149	149	149	149	149	149	149	149
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	2.021	1.999	1.994	1.989	1.979	1.927	1.932	1.931	1.969	1.970	1.921	1.942	1.945	1.952	1.945	1.938
p(t) =	3.712	3.735	3.670	3.665	3.657	3.657	3.642	3.674	3.653	3.634	3.637	3.631	3.605	3.601	3.600	3.592
H =	0.08929	0.08792	0.08782	0.08753	0.08701	0.08391	0.08424	0.08411	0.08760	0.08655	0.08362	0.08488	0.08513	0.08559	0.08514	0.08478
press corr for H =	0.004897	0.004631	0.004734	0.004712	0.004692	0.004530	0.004545	0.004542	0.004727	0.005281	0.005108	0.005181	0.005193	0.005220	0.005194	0.005170
p =	1.913	1.888	1.886	1.881	1.871	1.813	1.819	1.817	1.882	1.863	1.809	1.832	1.837	1.845	1.837	1.830
% RH =	51.5	50.5	51.4	51.3	51.2	49.6	49.9	49.4	51.5	51.3	49.7	50.5	51.0	51.3	51.0	51.0
% of gas from SupHtr	9.15	9.07	8.72	8.74	8.75	8.62	8.60	8.76	8.58	8.69	8.68	8.96	8.70	8.62	8.57	8.52
WL Exit Conditions:																
H =	0.10346	0.10177	0.10140	0.10108	0.10049	0.09678	0.09715	0.09716	0.10100	0.10057	0.09718	0.09894	0.09894	0.09937	0.09880	0.09833
p =	2.078	2.049	2.040	2.035	2.025	1.960	1.967	1.967	2.033	2.015	1.957	1.987	1.987	1.995	1.985	1.977
p(t) =	2.053	2.104	2.042	2.031	2.032	1.964	1.967	1.955	2.023	2.104	1.951	1.989	1.978	2.102	1.973	1.974
% RH =	101.2	97.4	99.9	100.2	99.7	99.8	100.0	100.6	100.5	95.8	100.3	99.9	100.5	94.9	100.6	100.2

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-45	NC-46	NC-47	NC-48	NC-49	NC-50	NC-51	NC-52	NC-53	NC-54	NC-55	NC-56	NC-57	NC-58	NC-59	NC-60
Nozzle Pressure, psig																
1	25	35	45	45	45	45	45	45	45	45	35	30	15	50	15	15
2	45	35	25	25	25	25	45	45	45	45	35	30	15	50	50	15
3	45	35	25	25	45	45	25	25	45	45	35	30	15	15	15	50
4	45	35	25	45	25	45	25	45	25	45	35	30	50	50	50	50
Nozzle Flow, gpm																
1	0.40	0.25	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.25	0.15	0.40	0.10	0.40	0.40
2	0.10	0.25	0.40	0.40	0.40	0.40	0.10	0.10	0.10	0.10	0.25	0.15	0.40	0.10	0.10	0.40
3	0.10	0.25	0.40	0.40	0.10	0.10	0.40	0.40	0.10	0.10	0.25	0.15	0.40	0.40	0.40	0.10
4	0.10	0.25	0.40	0.10	0.40	0.10	0.40	0.10	0.40	0.10	0.25	0.15	0.10	0.10	0.10	0.10
Total gal/min	0.70	1.00	1.30	1.00	1.00	0.70	1.00	0.70	0.70	0.40	1.00	0.60	1.30	0.70	1.00	1.00
No. of nozzles used	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	300	299	299	299	299	299	299	299	298	298	298	299	299	299	299	299
Flue Gas °F	432	415	418	419	423	424	426	432	435	437	428	429	418	425	407	407
Rx Inlet °F	130	130	130	131	131	130	131	130	130	130	130	131	131	130	130	131
Rx Exit °F	131	131	131	131	132	132	132	131	131	131	131	132	132	132	132	132
BH Inlet °F	134	134	135	135	135	135	135	134	134	135	135	135	135	135	135	135
BH Exit °F	149	149	149	149	150	150	150	149	150	149	150	150	150	150	150	150
BH Exit Wet Bulb °F	126	126	125	126	125	125	126	126	125	126	125	126	126	127	126	126
React Inlet Press., "H2O	−4.78	−4.64	−4.78	−4.78	−4.80	−4.78	−4.72	−4.79	−4.75	−4.74	−5.08	−4.93	−5.18	−4.71	−5.05	−4.96
WL Exit O2 vol %	13.6	13.4	13.0	13.3	13.4	13.6	14.0	14.1	13.8	14.0	13.2	13.5	12.6	13.9	13.4	13.6
BH Exit O2 vol %	13.7	13.7	13.4	13.7	13.8	14.0	14.0	14.3	14.1	14.3	13.5	13.8	12.9	14.2	13.8	13.9
WL Flow (std ft³/min)	488	486	486	486	482	484	511	511	509	511	495	490	497	476	466	467
Room Temp. °F	75	74	72	76	80	83	75	76	80	83	74	73	72	76	80	84
WL Inlet °F	280	280	281	277	278	279	280	280	281	278	280	280	280	281	280	281
WL Exit °F	126	126	127	126	125	125	127	126	125	125	125	127	128	127	127	127
WL Exit 2 °F	128	128	128	128	128	127	129	128	128	127	127	129	129	129	129	129
WL Exit Aft Bypass °F	130	130	130	130	130	130	130	129	129	129	129	130	130	130	130	130
Recycle FG °F	142	143	143	142	142	141	143	142	142	141	142	143	143	143	144	144
Sup Heat Air Temp °F	306	305	305	305	305	306	306	308	308	306	305	305	305	308	305	308
WL Exit Pressure, "H2O	−2.91	−2.74	−2.89	−2.88	−2.84	−2.84	−2.83	−2.91	−2.84	−2.82	−3.18	−3.04	−3.21	−2.81	−3.13	−3.01
Rel Hum WL Exit	89.4	90.4	86.4	85.7	84.6	84.8	85.8	90.1	88.8	89.7	84.9	86.2	85.8	88.8	86.3	86.6
Rel Hum BH Exit	51.1	51.3	50.3	50.9	50.4	50.4	50.1	51.4	50.6	50.9	49.8	49.5	50.0	52.1	50.1	49.3
Avg Rx Skin Temp °F	131.1	131.1	130.9	131.0	131.1	131.3	131.1	130.9	131.0	131.7	131.3	131.1	131.1	131.1	131.1	131.2
Barometric Pressure "Hg	28.85	28.85	28.85	28.85	28.85	28.85	28.73	28.73	28.73	28.73	28.87	28.87	28.87	28.87	28.87	28.87
BH Exit Conditions:																
Dry Bulb Temp.	149	149	149	149	150	150	150	149	150	149	150	150	150	150	150	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	1.974	1.976	1.942	1.965	1.947	1.947	1.971	1.960	1.955	1.961	1.918	1.983	1.968	2.013	1.975	1.994
p(t) =	3.652	3.650	3.642	3.649	3.658	3.665	3.697	3.648	3.660	3.654	3.669	3.710	3.697	3.677	3.700	3.716
H =	0.08671	0.08682	0.08484	0.08621	0.08511	0.08505	0.08641	0.08705	0.08555	0.08594	0.08339	0.08707	0.08622	0.08892	0.08864	0.08769
press corr for H =	0.004262	0.004264	0.004171	0.004238	0.004186	0.004184	0.004742	0.004770	0.004690	0.004710	0.004033	0.004209	0.004172	0.004289	0.004190	0.004240
p =	1.866	1.867	1.830	1.855	1.835	1.834	1.860	1.872	1.844	1.851	1.802	1.871	1.855	1.908	1.863	1.883
% RH =	51.1	51.1	50.2	50.8	50.2	50.0	50.3	51.3	50.4	50.7	49.1	50.4	50.2	51.8	50.4	50.7
% of gas from SupHtr	8.83	8.82	8.64	8.63	8.59	8.64	8.54	8.76	8.81	8.67	8.57	8.65	8.77	8.62	8.77	8.76
WL Exit Conditions:																
H =	0.09978	0.09990	0.09743	0.09899	0.09769	0.09768	0.09966	0.10064	0.09896	0.09925	0.09562	0.09991	0.09908	0.10201	0.09958	0.10075
p =	2.019	2.022	1.978	2.006	1.963	1.983	2.009	2.026	1.997	2.002	1.948	2.023	2.006	2.059	2.017	2.038
p(t) =	1.973	1.984	2.028	1.973	1.943	1.928	2.031	1.969	1.950	1.942	1.941	2.012	2.081	2.049	2.017	2.015
% RH =	102.3	101.9	97.6	101.7	102.1	102.8	98.9	101.8	102.4	103.1	100.4	100.5	96.5	100.5	100.0	101.1

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-61	NC-62	NC-63	NC-64	NC-65	NC-66	NC-67	NC-68	NC-69	NC-70	NC-71	NC-72	NC-73	NC-74	NC-75	NC-54.1
Nozzle Pressure, psig																
1	50	50	50	50	30	50	15	15	15	50	15	15	50	30	40	45
2	50	15	15	15	30	50	50	15	15	15	50	50	50	30	50	45
3	15	15	50	50	30	50	50	15	50	15	15	50	50	30	50	45
4	15	50	50	15	30	15	50	15	15	15	15	15	50	30	45	45
Nozzle Flow, gpm																
1	0.10	0.10	0.10	0.10	0.15	0.10	0.40	0.40	0.40	0.10	0.40	0.40	0.10	0.15	0.43	0.10
2	0.10	0.40	0.40	0.40	0.15	0.10	0.10	0.40	0.40	0.40	0.10	0.10	0.10	0.15	0.20	0.10
3	0.40	0.40	0.10	0.10	0.15	0.10	0.10	0.40	0.10	0.40	0.40	0.10	0.10	0.15	0.20	0.10
4	0.40	0.10	0.10	0.40	0.15	0.40	0.10	0.40	0.40	0.40	0.40	0.40	0.10	0.15	0.30	0.10
Total gal/min	1.00	1.00	0.70	1.00	0.60	0.70	0.70	1.60	1.30	1.30	1.30	1.00	0.40	0.60	1.13	0.40
No. of nozzles used	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	0	2	0	0	0	0	0	0	0	0	0	1	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	299	300	300	299	301	300	301	300	300	300	298	299	298	298	304	299
Flue Gas °F	434	434	437	434	417	434	427	402	414	413	417	425	443	430	418	441
Rx Inlet °F	130	130	130	130	130	130	130	133	131	131	135	133	132	132	132	131
Rx Exit °F	131	131	131	131	131	131	131	133	132	132	135	134	133	133	133	133
BH Inlet °F	135	135	135	135	135	134	134	136	135	135	139	138	137	137	136	137
BH Exit °F	149	149	150	150	149	149	149	150	150	150	151	151	150	150	150	150
BH Exit Wet Bulb °F	125	125	126	125	125	125	125	126	125	125	125	125	125	125	125	126
React Inlet Press., "H2O	−4.86	−4.93	−4.89	−4.98	−5.00	−4.88	−5.27	−5.40	−5.23	−5.18	−5.22	−5.10	−4.76	−5.12	−4.76	−5.54
WL Exit O2 vol %	13.8	13.4	13.6	13.4	13.5	13.7	13.5	12.1	12.9	13.0	13.1	13.6	14.3	13.3	13.7	14.5
BH Exit O2 vol %	13.9	13.5	14.0	13.8	13.7	13.8	13.9	12.6	13.3	13.1	13.4	13.9	14.5	13.7	14.3	14.5
WL Flow (std ft³/min)	535	539	537	544	538	539	540	537	539	542	538	538	538	538	534	535
Room Temp. °F	71	75	81	84	79	79	76	75	76	78	75	75	77	80	70	69
WL Inlet °F	280	280	280	280	279	281	281	281	280	279	280	279	280	280	280	282
WL Exit °F	127	126	128	128	126	126	126	134	130	132	134	129	126	126	126	126
WL Exit 2 °F	129	128	128	129	128	128	128	133	130	131	134	131	129	129	129	129
WL Exit Aft Bypass °F	129	129	129	130	129	129	129	133	131	131	135	133	131	131	130	131
Recycle FG °F	142	141	142	143	143	142	142	147	144	145	149	146	142	142	143	142
Sup Heat Air Temp °F	305	306	306	306	306	306	306	306	306	306	306	306	306	306	306	306
WL Exit Pressure, "H2O	−2.92	−3.07	−2.91	−2.95	−3.05	−2.89	−3.28	−3.34	−3.19	−3.12	−3.29	−3.14	−2.78	−3.14	−2.77	−3.69
Rel Hum WL Exit	88.2	87.8	88.0	87.4	88.9	88.9	88.0	81.4	84.9	84.7	75.4	77.4	83.8	83.8	83.6	85.4
Rel Hum BH Exit	50.0	49.8	50.9	50.8	50.8	50.4	50.3	49.2	50.8	50.7	48.0	47.9	49.1	48.5	50.0	49.6
Avg Rx Skin Temp °F	131.1	131.0	131.2	131.5	131.0	131.0	131.1	131.1	131.0	131.7	134.0	133.7	133.1	133.0	133.1	133.1
Barometric Pressure "Hg	28.90	28.90	28.90	28.90	28.84	28.84	28.84	28.84	28.84	28.84	28.54	28.54	28.54	28.54	28.90	28.85
BH Exit Conditions:																
Dry Bulb Temp.	149	149	150	150	149	149	149	150	150	150	151	151	150	150	150	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	1.938	1.929	1.961	1.954	1.955	1.945	1.942	1.959	1.958	1.955	1.944	1.941	1.952	1.943	1.958	1.989
p(t) =	3.645	3.635	3.684	3.665	3.639	3.633	3.634	3.699	3.662	3.678	3.803	3.763	3.698	3.708	3.686	3.712
H =	0.08458	0.08408	0.08589	0.08551	0.08584	0.08508	0.08488	0.08566	0.08572	0.08552	0.08448	0.08445	0.08529	0.08472	0.08556	0.08743
press corr for H =	0.003963	0.003940	0.004026	0.004011	0.004254	0.004221	0.004221	0.004271	0.004268	0.004257	0.005435	0.005423	0.005457	0.005431	0.004011	0.004321
p =	1.824	1.815	1.849	1.842	1.845	1.834	1.830	1.845	1.846	1.843	1.826	1.825	1.841	1.830	1.843	1.878
% RH =	50.0	49.9	50.5	50.3	50.7	50.5	50.4	49.9	50.4	50.1	48.0	48.5	49.8	49.4	50.0	50.6
% of gas from SupHtr	8.68	8.63	8.64	8.56	8.54	8.71	8.65	8.08	8.39	8.44	7.36	7.53	7.94	8.02	7.94	8.03
WL Exit Conditions:																
H =	0.09695	0.09634	0.09842	0.09790	0.09829	0.09780	0.09754	0.09783	0.09823	0.09808	0.09708	0.09720	0.09657	0.09801	0.09730	0.09976
p =	1.974	1.963	1.999	1.990	1.993	1.984	1.980	1.985	1.992	1.989	1.951	1.953	1.977	1.967	1.980	2.019
p(t) =	2.068	1.969	1.989	2.078	1.993	1.974	1.975	2.468	2.210	2.306	2.461	2.165	1.988	1.969	1.981	1.992
% RH =	95.5	99.7	100.5	95.8	100.0	100.5	100.2	80.5	90.1	86.2	79.3	90.2	99.4	99.9	101.0	101.3

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-55.1	NC-58.1	NC-76	NC-77	NC-78	NC-79	NC-80	NC-81	NC-82	NC-83	NC-84	NC-85	NC-86	NC-87	NC-88	NC-89
Nozzle Pressure, psig																
1	35	50	40	40				40	40		40			40	40	40
2	35	50	50		50			50		50		50		50	50	
3	35	15	50			50			50	50			50	50		50
4	35	50	45				45				45	45	45		45	45
Nozzle Flow, gpm																
1	0.25	0.10	0.43	0.43				0.43	0.43		0.43			0.43	0.43	0.43
2	0.25	0.10	0.20		0.20			0.20		0.20		0.20		0.20	0.20	
3	0.25	0.40	0.20			0.20			0.20	0.20			0.20	0.20		0.20
4	0.25	0.10	0.30				0.30				0.30	0.30	0.30		0.30	0.30
Total gal/min	1.00	0.70	1.13	0.43	0.20	0.20	0.30	0.63	0.63	0.40	0.73	0.50	0.50	0.83	0.93	0.93
No. of nozzles used	4	4	4	1	1	1	1	2	2	2	2	2	2	3	3	3
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	300	306	308	304	301	NA	307	303	298	299	301	301	301	301	308	306
Flue Gas °F	422	434	426	400	403	NA	409	455	443	448	433	430	423	437	436	436
Rx Inlet °F	132	132	131	139	147	NA	155	133	133	133	132	133	139	135	129	129
Rx Exit °F	133	133	133	139	146	NA	155	134	134	134	133	134	138	136	131	131
BH Inlet °F	137	137	137	142	150	NA	157	138	138	138	136	137	141	139	134	134
BH Exit °F	150	150	150	152	154	NA	159	150	150	150	150	150	152	151	149	149
BH Exit Wet Bulb °F	126	126	126	125	126	NA	126	125	125	125	125	126	125	125	125	125
React Inlet Press., "H2O	−5.74	−4.50	−4.65	−4.98	−5.10	NA	−4.88	−4.34	−4.44	−4.73	−4.55	−4.99	−4.78	−4.63	−4.63	−4.59
WL Exit O2 vol %	13.8	14.4	13.6	12.4	12.8	NA	11.9	12.5	13.2	12.8	12.0	12.7	12.2	13.4	13.0	12.8
BH Exit O2 vol %	13.7	14.3	13.8	12.5	12.9	NA	12.4	13.0	13.3	13.4	12.4	13.0	12.6	13.6	13.4	13.2
WL Flow (std ft³/min)	535	534	534	533	533	NA	540	540	540	540	537	536	536	536	540	540
Room Temp. °F	68	73	73	71	72	NA	81	78	79	80	82	82	85	85	77	76
WL Inlet °F	280	281	280	281	280	NA	280	281	281	280	280	279	279	280	280	280
WL Exit °F	126	126	126	142	152	NA	163	127	132	133	131	132	144	126	126	127
WL Exit 2 °F	130	129	130	140	149	NA	159	130	133	133	131	132	141	129	128	128
WL Exit Aft Bypass °F	131	131	131	140	149	NA	158	132	133	133	132	133	140	134	129	129
Recycle FG °F	143	142	143	154	159	NA	166	143	148	147	145	146	155	142	142	142
Sup Heat Air Temp °F	305	308	308	308	308	NA	305	305	305	305	305	305	305	305	305	305
WL Exit Pressure, "H2O	−3.81	−2.56	−2.68	−2.90	−3.12	NA	−2.95	−2.57	−2.60	−2.90	−2.73	−3.06	−2.86	−2.62	−2.72	−2.68
Rel Hum WL Exit	86.6	84.4	84.7	68.6	54.4	NA	43.1	79.8	80.5	80.0	81.3	80.9	68.4	73.3	88.2	88.0
Rel Hum BH Exit	49.5	50.3	50.3	46.0	44.6	NA	38.7	47.6	48.3	48.0	49.7	49.6	45.8	47.4	49.8	49.9
Avg Rx Skin Temp °F	133.1	133.1	133.1	136.6	142.9	NA	151.6	134.3	134.1	134.0	132.1	132.6	135.0	134.8	131.2	131.4
Barometric Pressure "Hg	28.85	28.90	28.90	28.70	28.64	NA	28.68	28.69	28.69	28.69	28.75	28.75	28.75	28.75	28.68	28.68
BH Exit Conditions:																
Dry Bulb Temp.	150	150	150	152	154	NA	159	150	150	150	150	150	152	151	149	149
lambda =	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08	NA	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08	2.38E+08
p(s) =	1.994	1.976	1.985	1.950	1.990	NA	1.987	1.934	1.955	1.947	1.948	1.964	1.946	1.927	1.928	1.928
p(t) =	3.718	3.695	3.692	3.870	4.123	NA	4.647	3.733	3.738	3.734	3.694	3.700	3.883	3.784	3.573	3.579
H =	0.08771	0.08670	0.08723	0.08464	0.08633	NA	0.08480	0.08414	0.08536	0.08486	0.08503	0.08598	0.08438	0.08355	0.08425	0.08419
press corr for H =	0.004340	0.004060	0.004086	0.004799	0.005185	NA	0.005002	0.004778	0.004849	0.004826	0.004584	0.004644	0.004582	0.004521	0.004804	0.004801
p =	1.883	1.864	1.874	1.828	1.860	NA	1.833	1.818	1.841	1.831	1.834	1.852	1.822	1.807	1.820	1.819
% RH =	50.7	50.5	50.8	47.2	45.1	NA	39.4	48.7	49.2	49.0	49.7	50.1	46.9	47.7	50.9	50.8
% of gas from SupHtr	8.05	7.83	7.82	5.79	2.81	NA	1.59	7.56	7.54	7.50	7.99	7.77	6.50	6.94	8.37	8.36
WL Exit Conditions:																
H =	0.10011	0.09848	0.09908	0.09493	0.09416	NA	0.09125	0.09619	0.09756	0.09695	0.09740	0.09826	0.09515	0.09464	0.09719	0.09711
p =	2.025	2.000	2.011	1.925	1.907	NA	1.858	1.946	1.970	1.959	1.971	1.966	1.932	1.923	1.963	1.961
p(t) =	2.000	1.983	1.991	3.060	3.881	NA	5.051	2.051	2.353	2.427	2.248	2.348	3.179	1.983	1.991	2.041
% RH =	101.2	100.9	101.0	62.9	49.1	NA	36.8	94.9	83.7	80.7	87.7	84.6	60.8	97.0	98.6	98.1

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	NC-90	RHM-01	RHM-02	RHM-03	RHM-04	RHM-05	RHM-06	RHM-07	RHM-08	RHM-09	RHM-10	RHM-11	RHM-12	RHM-13	RHM-14	RHM-15
Nozzle Pressure, psig																
1		40	40	30	20	40	30	20	40	30	20	40	30	20	40	30
2	50	50	40	30	20	40	30	20	40	30	20	40	30	20	40	30
3	50	50	40	30	20	40	30	20	40	30	20	40	30	20	40	30
4	45	45	40	30	20	40	30	20	40	30	20	40	30	20	40	30
Nozzle Flow, gpm																
1		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
2	0.20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
3	0.20	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
4	0.30	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total gal/min	0.70	1.13	1.13	1.13	1.13	0.70	0.70	0.70	0.60	0.60	0.60	0.55	0.55	0.55	0.50	0.50
No. of nozzles used	3	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	306	436	436	436	436	429	429	429	429	434	434	434	434	438	438	438
Flue Gas °F	441	427	421	408	403	404	404	402	415	403	394	408	408	395	397	397
Rx Inlet °F	130	129	129	129	131	130	130	131	130	131	135	134	134	136	131	129
Rx Exit °F	131															
BH Inlet °F	135	133	133	133	134	133	134	135	134	135	139	138	138	138	135	134
BH Exit °F	149	148	149	149	150	149	150	151	150	150	152	151	150	150	149	149
BH Exit Wet Bulb °F	126	124	125	125	124	125	125	126	125	125	125	125	125	126	125	125
React Inlet Press., "H2O	−4.87	−7.30	−7.43	−8.49	−9.24	−7.50	−8.14	−8.40	−7.00	−8.59	−8.98	−7.98	−8.30	−8.44	−8.09	−8.40
WL Exit O2 vol %	13.1	11.1	10.2	12.9	11.1	11.0	11.3	11.0	10.7	11.1	11.3	11.1	10.6	10.7	11.2	10.8
BH Exit O2 vol %	13.5	11.1	10.9	10.6	11.3	11.3	11.6	11.3	10.9	11.4	11.7	11.5	10.9	11.1	11.5	11.1
WL Flow (std ft³/min)	540	704	704	704	704	694	694	694	694	697	697	697	697	715	715	715
Room Temp. °F	78															
WL Inlet °F	280	283	280	279	280	282	280	280	281	282	281	280	275	280	280	280
WL Exit °F	128	125	126	128	134	127	127	134	126	128	137	126	127	133	126	127
WL Exit 2 °F	129	129	128	129	131	129	129	131	129	130	138	132	132	135	130	129
WL Exit Aft Bypass °F	129	128	128	128	131	128	129	131	129	130	135	133	133	135	130	129
Recycle FG °F	143															
Sup Heat Air Temp °F	305	497	499	502	509	509	509	509	509	508	509	497	489	325	433	528
WL Exit Pressure, "H2O	−2.93	−3.81	−3.94	−4.75	−5.55	−3.42	−4.31	−4.57	−3.19	−4.43	−4.91	−3.90	−4.22	−4.28	−3.73	−4.03
Rel Hum WL Exit	88.5															
Rel Hum BH Exit	51.0															
Avg Rx Skin Temp °F	131.6															
Barometric Pressure "Hg	28.68	29.08	29.08	29.08	29.10	29.17	29.20	29.20	29.22	29.16	29.16	29.13	29.14	28.97	29.00	29.00
BH Exit Conditions:																
Dry Bulb Temp.	149	148	149	149	150	149	150	151	150	150	152	151	150	150	149	149
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.39E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06
p(s) =	1.967	1.900	1.905	1.916	1.895	1.953	1.942	1.958	1.932	1.932	1.926	1.948	1.948	1.958	1.932	1.926
p(t) =	3.590	3.557	3.628	3.637	3.701	3.646	3.674	3.747	3.701	3.602	3.841	3.784	3.729	3.674	3.584	3.628
H =	0.08648	0.08208	0.08275	0.08334	0.08193	0.08548	0.08477	0.08549	0.08407	0.08410	0.08336	0.08476	0.08492	0.08571	0.08442	0.08398
press corr for H =	0.004935	0.003222	0.003236	0.003279	0.003172	0.002990	0.002865	0.002901	0.002746	0.003007	0.003004	0.003145	0.003112	0.003810	0.003618	0.003612
p =	1.861	1.787	1.789	1.800	1.774	1.840	1.826	1.840	1.813	1.814	1.801	1.827	1.830	1.846	1.821	1.813
% RH =	51.8	50.2	49.3	49.5	47.9	50.5	49.7	49.1	49.0	49.1	46.9	48.3	49.1	50.2	50.8	50.0
% of gas from SupHtr	8.18	4.31	4.53	4.55	4.32	4.28	4.09	4.12	4.19	3.89	3.51	3.51	3.54	6.31	4.57	3.93
WL Exit Conditions:																
H =	0.09954	0.08975	0.09007	0.09074	0.08894	0.09243	0.09137	0.09220	0.09061	0.09063	0.08951	0.09111	0.09126	0.09555	0.09226	0.09117
p =	2.003	1.857	1.863	1.875	1.843	1.911	1.894	1.909	1.882	1.878	1.858	1.885	1.888	1.953	1.897	1.878
p(t) =	2.113	1.953	1.974	2.098	2.446	2.018	2.051	2.433	2.001	2.107	2.660	2.001	2.051	2.420	2.007	2.040
% RH =	94.8	95.1	94.4	89.5	75.4	94.7	92.4	78.4	94.0	89.1	69.8	94.2	92.1	80.7	94.5	92.1

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	RHM-16	RHM-17	RHM-18	RHM-19	RHM-20	RHM-21	RHM-22	RHM-23	RHM-24	RHM-25	RHM-26	RHM-27	RHM-28	RHM-29	RHM-29.1	RHM-30
Nozzle Pressure, psig																
1	20	40	30	20	40	30	20	40	40	40	50	40	40	40	40	30
2	20	40	30	20	40	30	20	50	40	40	50	40	40	50	50	30
3	20	40	30	20	40	30	20	50	40	40	50			50	50	30
4	20	40	30	20	40	30	20	45	40	40	50	40	40	45	45	30
Nozzle Flow, gpm																
1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
3	0	0	0	0	0	0	0	0	0	0	0			0	0	0
4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total gal/min	0.50	0.45	0.45	0.45	0.60	0.60	0.60	1.13	1.03	0.60	0.60	0.50	0.40	1.13	1.13	0.60
No. of nozzles used	4	4	4	4	4	4	4	4	4	4	4	3	3	4	4	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0	0	0	0	0	20	22	21
Wash Water, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	2	2	2
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0	0	0	0	0	2	2	3
Rx Flow (scfm)	438	438	440	440	440	440	440	453	453	453	453	432	432	450	454	454
Flue Gas °F	394	400	401	395	398	398	392	453	453	453	454	453	452	387	399	380
Rx Inlet °F	132	132	132	137	131	131	134	128	128	129	128	130	132	128	129	130
Rx Exit °F																
BH Inlet °F	137	138	138	140	137	136	139	132	131	133	132	133	135	138	133	134
BH Exit °F	151	150	151	151	150	150	151	145	144	147	147	146	148	148	150	150
BH Exit Wet Bulb °F	125	125	125	125	125	125	125	121	121	123	122	123	124	125	124	126
React Inlet Press., "H2O	−8.69	−8.21	−8.81	−8.54	−8.13	−8.27	−8.38	−9.76	−9.89	−9.50	−9.37	−10.69	−10.72	−9.52	−8.80	−9.11
WL Exit O2 vol %	10.9	11.2	11.2	10.7	10.7	10.6	10.8	11.1	10.4	11.3	11.8	10.4	10.5	11.6	11.9	11.6
BH Exit O2 vol %	10.5	10.9	10.9	11.0	11.1	10.9	11.1	11.5	10.8	11.6	12.0	10.9	10.9	11.8	12.0	11.8
WL Flow (std ft³/min)	715	715	714	714	714	714	714	863	863	863	863	842	842	707	712	712
Room Temp. °F																
WL Inlet °F	280	280	281	281	279	281	280	280	280	283	282	280	280	276	281	282
WL Exit °F	133	126	130	134	127	128	134	124	124	127	125	131	132	126	123	126
WL Exit 2 °F	133	131	131	136	130	130	134	127	127	129	127	131	132	129	126	128
WL Exit Aft Bypass °F	133	132	131	136	130	130	134	127	126	129	127	130	132	128	128	129
Recycle FG °F																
Sup Heat Air Temp °F	506	489	489	403	481	498	484	407	408	500	509	388	397	597	647	648
WL Exit Pressure, "H2O	−4.40	−3.79	−4.48	−4.29	−3.75	−3.97	−4.03	−5.44	−5.55	−5.10	−4.80	−6.56	−6.53	−3.51	−3.34	−3.38
Rel Hum WL Exit																
Rel Hum BH Exit																
Avg Rx Skin Temp °F																
Barometric Pressure "Hg	29.00	29.00	29.00	29.05	29.13	29.15	29.15	28.99	29.03	29.05	29.05	29.15	29.15	29.12	29.00	28.95
BH Exit Conditions:																
Dry Bulb Temp.	151	150	151	151	150	150	151	145	144	147	147	146	148	148	150	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+03	2.38E+06	2.38E+06	2.38E+06	2.39E+06	2.39E+06	2.39E+06	2.39E+06	2.39E+06	2.39E+06	2.38E+06	2.39E+06	2.38E+06
p(s) =	1.926	1.905	1.948	1.948	1.937	1.942	1.932	1.706	1.721	1.809	1.789	1.824	1.864	1.916	1.874	1.974
p(l) =	3.747	3.720	3.757	3.813	3.692	3.692	3.794	3.266	3.209	3.460	3.443	3.349	3.495	3.478	3.665	3.683
H =	0.08363	0.08249	0.08484	0.08468	0.08441	0.08472	0.08380	0.07236	0.07337	0.07763	0.07654	0.07884	0.08074	0.08382	0.08084	0.08663
press corr for H =	0.003617	0.003557	0.003672	0.003469	0.003125	0.003061	0.003042	0.003146	0.003049	0.003169	0.003122	0.002864	0.002947	0.003146	0.003493	0.003946
p =	1.807	1.785	1.830	1.826	1.820	1.826	1.809	1.560	1.609	1.692	1.671	1.714	1.751	1.809	1.754	1.864
% RH =	48.2	48.0	48.7	47.9	49.3	49.5	47.7	48.7	50.2	48.9	48.5	51.2	50.1	52.0	47.9	50.6
% of gas from SupHtr	3.79	3.62	3.70	4.12	3.86	3.78	3.51	4.86	4.81	3.92	3.95	5.29	4.95	2.11	3.17	3.04
WL Exit Conditions:																
H =	0.09069	0.08927	0.09182	0.09194	0.09105	0.09123	0.09000	0.07937	0.08027	0.08409	0.08294	0.08627	0.08805	0.08884	0.08709	0.09341
p =	1.869	1.844	1.891	1.894	1.884	1.888	1.868	1.662	1.681	1.752	1.731	1.798	1.830	1.843	1.804	1.914
p(l) =	2.389	1.996	2.182	2.485	2.023	2.079	2.433	1.854	1.900	2.018	1.905	2.247	2.357	1.990	1.804	1.964
% RH =	78.2	92.4	86.7	76.2	93.1	90.8	76.7	89.6	88.5	86.9	90.9	80.0	77.6	92.6	100.0	97.5

TABLE 11 (Continued)

HUMIDIFICATION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	RHM-31	RHM-32	RHM-33	RHM-34	RHM-35	RHM-36	RHM-36.1	RHM-37	RHM-38
Nozzle Pressure, psig									
1	30	40	40	40	50	40	40	40	40
2	30	50	40	40	50	40	40	40	40
3		50	40	40	50				40
4	30	45	40	40	50	40	40	40	40
Nozzle Flow, gpm									
1	0	0	0	0	0	0	0	0	0
2	0	0	0	0	0	0	0	0	0
3	.	0	0	0	0				0
4	0	0	0	0	0	0	0	0	0
Total gal/min	0.60	1.13	1.13	0.60	0.60	0.50	0.50	0.40	0.55
No. of nozzles used	3	4	4	4	4	3	3	3	4
Ash Feed Rate, lb/hr	0	0	0	0	0	0	0	0	0
Wash Water, gpm	0	0	0	0	0	0	0	0	0
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0
Rx Flow (scfm)	308	488	488	488	488	500	501	501	501
Flue Gas °F	404	467	462	474	474	459	442	442	370
Rx Inlet °F	131	129	130	134	133	135	138	138	130
Rx Exit °F									
BH Inlet °F	134	133	133	135	138	138	140	140	134
BH Exit °F	149	145	145	150	149	150	150	150	149
BH Exit Wet Bulb °F	125	123	122	125	125	124	123	123	125
React Inlet Press., "H2O	−4.37	−10.91	−11.21	−11.40	−10.29	−11.24	−11.11	−11.11	−8.47
WL Exit O2 vol %	12.0	10.2	10.1	10.8	11.2	10.0	11.1	10.9	10.5
BH Exit O2 vol %	12.3	10.5	10.4	11.1	11.5	10.3	11.2	11.1	10.8
WL Flow (std ft³/min)	512	1035	1035	1035	1035	1050	985	985	702
Room Temp. °F									
WL Inlet °F	281	273	271	281	273	275	280	280	278
WL Exit °F	127	128	128	134	127	136	138	139	126
WL Exit 2 °F	130	128	129	134	131	136	140	140	129
WL Exit Aft Bypass °F	130	129	129	134	131	135	139	139	129
Recycle FG °F									
Sup Heat Air Temp °F	291	394	408	527	485	449	325	315	597
WL Exit Pressure, "H2O	−2.14	−5.92	−6.46	−6.70	−4.79	−6.21	−5.48	−5.51	−3.37
Rel Hum WL Exit									
Rel Hum BH Exit									
Avg Rx Skin Temp °F									
Barometric Pressure "Hg	29.16	29.10	29.10	29.13	29.13	29.00	28.80	28.80	28.82
BH Exit Conditions:									
Dry Bulb Temp.	149	145	145	150	149	150	150	150	149
lambda =	2.38E+06	2.39E+06	2.39E+06	2.38E+06	2.38E+06	2.39E+06	2.39E+06	2.39E+06	2.38E+06
p(s) =	1.926	1.809	1.769	1.932	1.937	1.860	1.829	1.839	1.905
p(t) =	3.646	3.250	3.274	3.683	3.619	3.674	3.729	3.656	3.601
H =	0.08392	0.07829	0.07593	0.08413	0.08462	0.08051	0.07799	0.07879	0.08283
press corr for H =	0.002915	0.003015	0.002936	0.003177	0.003167	0.003526	0.004159	0.004188	0.004257
p =	1.810	1.704	1.659	1.815	1.825	1.748	1.702	1.717	1.793
% RH =	49.6	52.4	50.7	49.3	50.4	47.6	45.6	47.0	49.8
% of gas from SupHtr	9.91	4.70	4.44	3.67	3.73	3.89	5.52	5.58	3.28
WL Exit Conditions:									
H =	0.09639	0.08532	0.08253	0.09063	0.09119	0.08744	0.08695	0.08788	0.09003
p =	1.982	1.777	1.726	1.875	1.886	1.810	1.789	1.805	1.846
p(t) =	2.012	2.067	2.084	2.427	2.051	2.571	2.702	2.766	1.960
% RH =	96.5	86.0	82.8	77.3	92.0	70.4	66.2	65.3	93.2

TABLE 12

PARTICULATE COLLECTION EFFICIENCY TESTS USING SECOND GENERATION CONTACTOR

Run No.	PCE-01	PCE-02	PCE-03	PCE-04	PCE-05	PCE-06	PCE-07	PCE-08	PCE-09	PCE-10	PCE-11	PCE-12	PCE-13	PCE-14	RHM-29.1	RHM-30
Nozzle Air Pressure, psig																
1	40	30	15	50	15	50	15	15	50	50	30	15	50	50	40	30
2	50	30	15	50	15	50	50	50	15	15	30	50	15	15	50	30
3	50	30	15	50	50	15	50	15	50	15	30	50	50	15	50	30
4	45	30	15	50	50	15	50	15	50	15	30	15	15	50	45	30
Nozzle Water Flow, gpm																
1	0.43	0.15	0.40	0.10	0.40	0.10	0.40	0.40	0.10	0.10	0.15	0.40	0.10	0.10	0.43	0.15
2	0.2	0.15	0.40	0.10	0.40	0.10	0.10	0.10	0.40	0.40	0.15	0.10	0.40	0.40	0.2	0.15
3	0.2	0.15	0.40	0.10	0.10	0.40	0.10	0.40	0.10	0.40	0.15	0.10	0.10	0.40	0.2	0.15
4	0.3	0.15	0.40	0.10	0.10	0.40	0.10	0.40	0.10	0.40	0.15	0.40	0.40	0.10	0.3	0.15
Total gal/min	1.13	0.60	1.60	0.40	1.00	1.00	0.70	1.30	0.70	1.30	0.60	1.00	1.00	1.00	1.13	0.60
No. of nozzles used	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4	4
Ash Feed Rate, lb/hr	20.1	20.0	20.0	20.0	20.1	20.3	20.0	20.1	20.0	20.1	20.0	17.8	20.0	19.8	21.7	21.2
Wash Water, gpm	2	2	2	2	2	2	2	2	2	2	1	2	2	2	1.75	1.8
Mist Elim Wash, gpm	0	0	0	0	0	0	0	0	0.6	0.6	0.6	0.6	0.6	0.9	2	3
Rx Flow (scfm)	304	306	305	303	303	304	303	301	304	305	297	298	308	308	454	454
Flue Gas °F	458	435	429	450	429	432	434	437	451	440	436	430	424	400	399	380
Rx Inlet °F	131	133	133	132	132	132	131	132	132	132	132	132	132	132	129	130
Rx Exit °F	133	134	134	133	133	133	133	133	133	133	133	134	134	133		
BH Inlet °F	137	138	138	137	137	137	137	137	137	137	137	137	137	137	133	134
BH Exit °F	150	150	150	150	150	150	150	150	150	150	150	150	150	149	150	150
BH Exit Wet Bulb °F	126	125	125	125	125	125	125	124	126	126	126	126	125	125	124	126
React Inlet Press., "H2O	−4.62	−4.29	−6.95	−3.53	−4.13	−4.96	−6.68	−6.33	−3.66	−4.48	−4.93	−6.80	−5.81	−6.39	−8.80	−9.11
WL Exit O2 vol %	15.0	14.7	14.0	15.9	14.6	14.9	14.6	12.6	14.9	14.4	14.6	14.8	14.6	14.7	11.9	11.6
BH Exit O2 vol %	15.2	14.8	14.4	15.6	14.8	15.0	14.7	12.8	15.1	14.7	14.7	14.9	14.8	14.9	12.0	11.8
WL Flow (std ft³/min)	534	534	533	532	533	533	532	533	535	534	534	534	535	535	712	712
Room Temp. °F	67.7	75.0	77.7	69.6	71.6	74.6	77.9	76.4	74.4	79.0	77.0	81.4	78.1	81.5		
WL Inlet °F	278	278	282	281	281	279	280	280	277	282	281	275	281	280	281	282
WL Exit °F	126	126	124	126	126	126	126	126	127	127	127	127	127	126	123	126
WL Exit 2 °F	130	129	129	127	128	128	128	130	130	130	130	130	130	129	126	128
WL Exit Aft Bypass °F	131	132	132	130	131	131	131	132	132	132	131	132	132	131	126	129
Recycle FG °F	143	142	142	140	142	142	142	143	142	144	142	144	144	144		
Sup Heat Air Temp °F	306	306	306	305	305	305	305	305	305	305	306	306	305	305	647	648
WL Exit Pressure, "H2O	−2.54	−2.25	−4.90	−1.46	−2.03	−2.89	−4.58	−4.40	−1.64	−2.37	−2.64	−4.58	−3.51	−3.76	−3.34	−3.38
Rel Hum WL Exit	85.9	78.9	78.2	84.3	83.4	83.4	84.5	79.4	84.8	82.6	82.4	81.0	80.9	79.8		
Rel Hum BH Exit	49.8	47.6	46.8	50.0	50.4	49.6	49.4	46.8	52.2	51.4	50.8	50.0	50.7	50.0		
Avg Rx Skin Temp °F	133.1	133.8	133.9	132.9	133.1	133.1	133.0	133.0	133.4	133.6	133.8	134.2	134.0	134.0		
Barometic Pressure "Hg	28.90	28.90	28.70	28.70	28.70	28.72	28.74	28.64	28.76	28.76	28.76	28.76	28.71	28.71	29.00	28.95
BH Exit Conditions:																
Dry Bulb Temp.	150	150	150	150	150	150	150	150	150	150	150	150	150	149	150	150
lambda =	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.38E+06	2.39E+06	2.38E+06
p(s) =	1.994	1.931	1.907	1.957	1.955	1.947	1.951	1.904	2.003	1.985	1.964	1.966	1.955	1.934	1.874	1.974
p(l) =	3.713	3.727	3.733	3.682	3.675	3.687	3.689	3.726	3.659	3.675	3.671	3.691	3.663	3.653	3.665	3.683
H =	0.08772	0.08393	0.08251	0.08561	0.08540	0.08504	0.08528	0.08237	0.08840	0.08728	0.08603	0.08609	0.08553	0.08437	0.08084	0.08663
press corr for H =	0.004111	0.003933	0.004700	0.004794	0.004799	0.004712	0.004678	0.004915	0.004696	0.004656	0.004600	0.004644	0.004792	0.004739	0.003493	0.003946
p =	1.883	1.812	1.787	1.845	1.843	1.835	1.839	1.785	1.897	1.876	1.853	1.854	1.844	1.822	1.754	1.864
% RH =	50.7	48.6	47.9	50.1	50.2	49.8	49.8	47.9	51.8	51.0	50.5	50.2	50.3	49.9	47.9	50.6
% of gas from SupHtr	8.03	7.39	7.39	7.82	7.43	7.48	7.54	7.81	7.48	7.51	7.36	7.39	7.26	7.33	3.17	3.04
WL Exit Conditions:																
H =	0.09984	0.09466	0.09417	0.09807	0.09754	0.09701	0.09728	0.09466	0.10063	0.09936	0.09763	0.09798	0.09739	0.09616	0.08709	0.09341
p =	2.024	1.937	1.911	1.980	1.970	1.962	1.968	1.916	2.028	2.006	1.979	1.981	1.968	1.946	1.804	1.914
p(l) =	2.005	1.963	1.887	1.961	1.990	1.978	1.960	1.982	2.057	2.065	2.024	2.022	2.026	1.959	1.804	1.964
% RH =	101.0	98.7	101.2	101.0	99.0	99.2	100.4	96.7	98.6	97.2	97.6	98.0	97.2	99.4	100.0	97.5
wt % Fly Ash Capture =	96.7	94.1	93.7	96.5	95.8	95.5	96.2	95.2	94.4	91.8	92.4	90.5	93.8	90.1	82.8	84.5

TABLE 13

HUMIDIFICATION TESTS OF AS-RECEIVED VENTURI CONTACTOR

Flue Gas Temp, °F	Flue Gas Flow, scfm	Contactor ΔP "H_2O	Contactor Exit Wet Bulb, °F	H_2O GPM at Throat	Approach to Saturation, °F
280	806	5	130.6	5	25.4
280	543	5	128.2	5	14.8
274	478	5	129.0	5	13.9

TABLE 14

HUMIDIFICATION TESTS OF ORIGINAL VENTURI CONTACTOR WITH UPSTREAM HYDRAULIC NOZZLES

Flue Gas Temp., °F	Flue Gas Flow, scfm	Contactor ΔP, "H_2O	Contactor Wet Bulb, °F In	Contactor Wet Bulb, °F Out	H_2O Flow, gpm at Throat	H_2O Flow, gpm Upstream Nozzles	Approach to Saturation, °F
283	473	0	125	120	2.2	1.4	14
279	473	6	123	119	2.2	1.4	11
281	473	4	123	119	2.2	1.8	10
280	473	4	122	119	3.2	1.4	12
278	570	4	125	121	2.7	1.6	10
280	709	4	125	122	2.2	1.4	28
279	476	4	124	119	2.2	1.8	0.6
282	476	4	128	122	3.2	1.8	2.9
283	663	6	126	121	3.2	1.8	8.0
281	348	5	127	121	2.7	1.6	1.4
281	225	6	127	121	2.2	1.8	0.2
282	225	4	127	120	3.2	1.8	0.5
280	472	4	125	122	3.2	1.8	2.4
281	665	5	125	123	3.2	1.8	8.9
280	207	5	127	120	3.2	1.8	3.5
281	681	6	126	123	3.2	2.0	7.1
280	348	5	126	121	2.2	1.8	0.7
280	207	5	125	118	3.2	2.0	4.7
280	695	4	124	122	3.2	1.8	13
279	260	6	127	100	3.2	1.8	3.6
283	681	4	127	125	2.6	2.0	13
280	677	5	126	123	2.6	2.0	7.3
282	471	4	126	121	2.6	2.0	3.6
283	459	5	125	121	2.6	2.0	3.2
280	470	6	130	125	2.2	1.4	8.3
280	470	6	140	135	2.2	1.4	8.5
280	470	6	134	130	3.2	1.4	8.2
279	470	6	141	136	3.2	1.4	5.9
281	470	7	126	126	3.2	1.4	16
280	471	6	124	124	3.2	1.4	16
280	471	7	124	107	3.2	1.4	8.9
282	446	6	127	123	2.6	2.0	2.2

TABLE 15

HUMIDIFICATION TESTS OF MODIFIED VENTURI CONTACTOR

Flue Gas Temp., °F	Flue Gas Flow, scfm	Contactor ΔP, "H_2O	Contactor Wet Bulb, °F In	Contactor Wet Bulb, °F Out	H_2O Flow, gpm at Throat	H_2O Flow, gpm Other H_2O Addition	Approach to Saturation, °F
288	642	4	95	117	2.6	0	38
289	642	7	125	122	3.4	0	10
279	642	8	125	121	3.4	0	6.9
283	635	8	127	122	4.2	0	4.3
284	635	8	124	122	2.6	0	13
280	635	8	124	122	5.0	0	3.6
281	635	5	125	123	2.6	0	29
284	635	5	126	123	3.4	0	20
283	635	5	127	124	4.2	0	11
281	635	5	127	124	5.0	0	7.1
281	635	6	126	123	2.6	0	23
281	635	6	125	123	3.4	0	13
280	635	6	125	123	4.2	0	8.1
280	635	6	127	124	5.0	0	5.0
280	635	7	127	124	2.6	0	15
280	635	7	126	123	4.2	0	6.3
280	635	7	127	124	5.0	0	4.0
280	635	8	127	123	3.4	1.5*	2.9
280	635	8	126	123	3.4	1.75*	3.2
282	643	8	126	121	3.0	2.0*	3.8
282	643	7	126	122	3.4	1.5*	5.7
283	643	7	127	123	3.4	1.75*	4.6
284	643	7	126	123	3.0	2.0*	4.6
280	467	6	125	122	5.0	0	3.7
281	467	6	125	122	4.2	0	5.0
282	467	6	127	123	3.4	0	7.6
280	467	6	125	122	2.6	0	13
280	467	7	124	121	2.6	0	9.3
280	467	7	125	121	3.4	0	5.3
280	467	7	125	121	4.2	0	3.5
283	467	7	126	121	5.0	0	3.6
279	465	8	125	119	3.4	1.5*	2.0
282	465	8	122	111	3.4	1.75*	2.1
282	465	8	122	113	3.0	2.0*	2.0
283	465	6	126	123	3.4	1.75*	2.7
283	465	6	126	123	3.0	2.0*	3.0
282	465	7	126	123	3.4	1.5*	3.1
281	465	7	125	122	3.4	1.75*	3.3
280	465	7	125	122	3.0	2.0*	3.2
282	465	6	127	123	3.4	1.5*	3.5

*Downstream of Throat
**Upstream of Throat

TABLE 15 (Continued)

HUMIDIFICATION TESTS OF MODIFIED VENTURI CONTACTOR

Flue Gas Temp., °F	Flue Gas Flow, scfm	Contactor ΔP, "H_2O	Contactor Wet Bulb, °F In	Contactor Wet Bulb, °F Out	H_2O Flow, gpm at Throat	H_2O Flow, gpm Other H_2O Addition	Approach to Saturation, °F
280	461	8	128	123	2.6	0	6.4
280	461	8	127	124	3.4	0	3.5
280	461	8	126	123	4.2	0	2.2
280	461	8	126	122	5.0	0	2.2
282	455	9	125	121	2.6	0	6.8
282	455	9	126	121	3.4	0	3.9
280	455	9	127	121	4.2	0	2.5
280	455	9	126	122	5.0	0	1.9
282	539	6	128	125	3.4	1.5**	4.0
283	539	6	127	124	3.4	1.75**	2.5
281	539	6	128	125	3.0	2.0**	1.8
279	511	7	128	125	3.4	1.5**	2.4
280	511	7	128	124	3.4	1.75**	2.0
281	511	7	127	124	3.0	2.0**	1.4
280	483	8	126	124	3.0	2.0**	1.7
281	483	8	126	123	3.4	1.75**	2.0
281	483	8	126	123	3.4	1.5**	2.0
282	579	8		128	4.0	0	2.1
286	579	8		128	4.0	0.3*	0.2
287	579	8		129	3.8	0.3*	1.0
287	579	8		129	3.8	0.3*	1.6
297	579	8		129	4.8	0.3*	2.9
284	579	8		128	4.0	0.3*	1.2
280	579	8		127	4.0	0.3*	0.8
288	578	8		128	4.0	0.3*	0.4
290	578	8		129	4.0	0.3*	0
280	575	8		128	4.0	0.3*	0.6
282	575	8		130	4.0	0.3*	1.0
283	575	8		129	4.0	0.3*	0.2
291	579	8		130	4.0	0.3*	0.2
291	579	8		128	4.0	0.3*	0.8
291	579	8		129	4.0	0.3*	0.4
291	583	8		130	4.0	0.3*	0.6
290	583	8		128	4.0	0.3*	0.4
295	583	8		129	4.0	0.3*	0.4
290	578	8		129	4.0	0.3*	0.8
291	578	8		129	4.0	0.3*	0.2
281	588	8		131	4.0	0.3*	0.8
284	588	8		129	4.0	0.3*	0.2

* Downstream of Throat
** Upstream of Throat

TABLE 16

HUMIDIFICATION TESTS OF MODIFIED VENTURI CONTACTOR WITH SIMULATED MIST CARRY-OVER

Flue Gas Temp., °F	Flue Gas Flow, scfm	Contactor ΔP, "H_2O	Wet Bulb at Contactor Outlet , °F	H_2O Flow, gpm		Approach to Saturation, °F
				At Throat	Two-Fluid Nozzle at Separator Exit	
295	666	6	128	5.0	0	5.2
294	666	6	128	5.0	0.025	2.6
292	666	6	128	5.0	0.05	0
292	666	6	127	5.0	0	5.6
291	666	6	127	5.0	0.075	0.4
289	666	6	127	5.0	0.0375	0
294	572	8	128	5.0	0.0047	0
281	576	8	128	5.0	0.0041	0.6
282	576	8	128	5.0	0.0041	0

TABLE 17

HUMIDIFICATION TESTS OF VENTURI CONTACTOR WITH STEAM INJECTION AT SEPARATOR EXIT

Flue Gas Temp., °F	Flue Gas Flow, scfm	Contactor ΔP, "H_2O	Wet Bulb at Contactor Outlet , °F	H_2O Flow at Throat, gpm	Steam Injection, lb/hr	Approach to Saturation, °F
297	650	6	129	5.0	5.33	3.8
293	650	6	130	5.0	6.52	3.6
291	650	6	131	5.0	9.3	2.6
291	650	6	131	5.0	11.0	3.0
291	650	6	131	5.0	12.4	2.4
291	650	6	132	5.0	15.5	2.8
292	650	6	132	5.0	14.9	2.4
292	650	6	133	5.0	17.2	1.8
292	650	6	134	5.0	18.6	2.0
293	650	6	132	5.0	13.1	2.4
293	650	6	133	5.0	17.2	1.8
293	650	6	134	5.0	18.6	1.6
293	650	6	134	5.0	20.3	1.8
293	650	6	133	5.0	15.7	2.2
292	567	8	130	5.0		1.4
292	567	8	130	5.0	3.1	0.8
292	567	8	129	5.0	4.7	2.2
292	567	8	131	5.0	4.7	1.0
292	567	8	130	5.0	4.8	1.0
293	567	8	130	5.0	4.9	1.2
292	567	8	131	5.0	8.5	1.4
292	567	8	129	5.0	10.2	2.2
293	567	8	130	5.0	10.2	1.0
292	567	8	133	5.0	16.8	0.6
293	567	8	135	5.0	19.6	0.8
287	585	8	138	5.0	20	0.2
286	580	8	136	5.0	20	0.9
282	575	8	138	5.0	29	0.4
281	575	8	138	5.0	30	0
284	575	8	137	5.0	29	0
287	573	8	134	5.0	16	0
288	573	8	134	5.0	18	0
283	577	8	133	5.0	18	0
281	577	8	131	5.0	19	0.4
282	589	8	134	5.0	16	0

TABLE 18

HUMIDIFICATION TESTS OF ALTERNATE CONTACTOR DESIGN WITHOUT VENTURI

Flue Gas Temp., °F	Flue Gas Flow, scfm	H_2O Flow, GPM (a)	Approach to Saturation, °F
280	575	1.1	0.8
280	575	1.1	0.8
280	575	0.9	1.6
280	575	1.1	0.2
280	575	1.1	0
290	575	1.1	1.0
290	575	0.9	0.2
294	583	1.1	0
291	583	1.1	1.0
290	583	1.1	0.8
291	573	1.1	0.8
289	573	1.1	0.4
290	573	1.1	1.3
293	573	1.1	0.2
290	575	1.1	0.2
292	587	1.0	0.2
292	587	1.0	0.8

Note: ΔP <2" H_2O

(a) One hydraulic nozzle and five two-fluid nozzles (different locations and flow directions).

TABLE 19

SUMMARY OF PARTICULATE COLLECTION EFFICIENCY TESTS OF THIRD GENERATION (VENTURI) CONTACTOR

Test No.	Flue Gas Flow, acfm	Flue Gas Temp., °F	Fly Ash Load, gr/scf	Contactor Design	Water Flow, gpm	Contactor Pressure Drop, Inches, WC	Fly Ash Collection Efficiency, % by Mass
1A	1024	284	3.52	As Received(1)	5.0 (3.2 at throat, 1.8 upstream)	5.0	99.1
1B	380	280	3.37	As Received(1)	5.0 (3.2 at throat, 1.8 upstream)	5.0	99.6
2A	1020	305	3.55	Modified(2)	5.0 (All at throat)	8.0	99.8
2C	395	305	3.22	Modified(2)	5.0 (All at throat)	8.0	99.9

(1) The Fisher-Klosterman design with the venturi throat and the cyclonic separator close coupled. Some water was added upstream of the venturi.

(2) The modified version with a six foot transition between the venturi and the cyclonic separator.

TABLE 20

TEST CONDITIONS AND RESULTS, INITIAL RECYCLE TESTS OF THIRD GENERATION (VENTURI) CONTACTOR

Test	23	25–1	25–2	25–3	25–5	25–6	23B
Run Time, hr	275	112	112	112	112	112	15
Additive	None	None	None	None	None	None	NaCl
Na/Ca Mol Ratio	0.00	0.00	0.00	0.00	0.00	0.00	0.02
Sorbent Data							
Fresh Ca/S Mole Ratio (a)	1.38	1.32	1.34	1.40	1.33	1.33	1.30
Fresh Feedrate, lb/hr (b)	7.14	6.82	7.00	7.23	6.90	6.90	6.83
Recycle Feedrate, lb/hr	53.77	53.98	53.95	53.34	53.85	53.85	53.39
Recycle Ratio, lb recycle/lb fresh lime	7.53	7.91	7.71	7.38	7.80	7.80	7.82
Recycle Ratio, dry basis	6.17	6.48	6.31	6.04	6.39	6.39	6.40
Recycle Available Ca/S, Mol Ratio (a)	2.03	–	–	–	–	–	–
Total Available Ca/S, Mol Ratio (a)	3.41	–	–	–	–	–	–
Water Addition, lb/hr	6.74	6.01	6.01	5.94	6.00	6.00	5.95
lb Water/lb Recycle Sorbent	0.14	0.12	0.12	0.12	0.12	0.12	0.12
Duct Flue Gas Conditions							
In–Duct Residence Time, s	2.7	2.7	2.7	2.7	2.7	2.7	2.7
Duct Inlet SO_2 Content, ppmv–dry	1487	1493	1502	1488	1492	1496	1509
Approach to Saturation, °F							
Contactor Exit	2–4	<1	<1	1	<1	0	–
Duct Exit	4	4	4	4	3	3	4
Baghouse Exit	22	23	23	22	24	23	23
Solids Loading, gr/scf	20.9	20.9	20.9	20.8	20.8	20.8	20.7
Contactor Inlet Temp, °F	295	291	293	290	282	282	293
Contactor Exit Temp, °F	132	130	129	130	129	128	130
Duct Exit Temp, °F	133	132	132	132	131	131	132
Baghouse Exit Temp, °F	150	150	150	149	151	150	150
Baghouse Exit Wet Bulb, °F	128	127	127	127	127	127	127
Duct Inlet Flue Gas Flow, scfm	340	340	340	340	340	340	340
SO_2 Removal, %							
In–Duct	83	81	81	85	84	86	76
System (Duct + Baghouse)	92	85	84	92	89	89	89
Sorbent Utilization, %							
Steady State	67	64	63	65	67	67	69
Ash Analysis	67	62	63	–	62	–	–

(a) Includes all calcium in fresh and recycle sorbents not tied up with sulfur (e.g., $Ca(OH)_2$ and $CaCO_3$).
(b) Fresh feed was Mississippi hydrated lime.

TABLE 21

SUMMARY OF RECYCLE TEST RESULTS, INITIAL TESTING OF THIRD GENERATION (VENTURI) CONTACTOR

Test	13	23	25–1	25–2	25–3	25–5	25–6	23B
Sorbent/Duct Conditions								
Fresh Ca/S Mole Ratio	1.21	1.38	1.32	1.34	1.40	1.33	1.33	1.30
Recycle Ratio, dry basis	6.9	6.2	6.5	6.3	6.0	6.4	6.4	6.4
Na/Ca Mol Ratio	0	0	0	0	0	0	0	0.02
Approach to Saturation, °F								
Contactor Exit	<1	2–4	<1	<1	1	<1	0	–
Duct Exit	4	4	4	4	4	3	3	4
Baghouse Exit	23	22	23	23	22	24	23	23
Nozzle Conditions								
Venturi Throat Nozzles, gpm	(a)	5	4	4	4	5	5	5
Venturi Press., inches water	–	8	8	8	8	8	8	8
Atomization Nozzle 1 (b)								
Air Press., psig	–	–	47	47	40	–	–	–
Water Flow, gpm	–	0	0.15	0.15	0.15	0	0	0
Atomization Nozzle 2 (b)								
Air Press., psig	–	–	45	46	40	–	–	–
Water Flow, gpm	–	0	0.15	0.15	0.15	0	0	0
Atomization Nozzle 3 (c)								
Air Press., psig	–	–	–	–	–	45	90	–
Water Flow, gph	–	0	0	0	0	0.15	0.54	0
SO_2 Removal, %								
In–Duct	87	83	81	81	85	84	86	76
System (Duct + Baghouse)	90	92	85	84	92	89	89	89
Sorbent Utilization, % (d)	75	67	64	63	65	67	67	69

(a) Run 13 was made with the second generation contactor.
(b) Located upstream of the venturi throat.
(c) Located downstream of the cyclonic separator, and upstream of sorbent injection.
(d) Based on flue gas analysis and fresh sorbent feed rate/composition, assuming steady state.

TABLE 22

TEST CONDITIONS AND RESULTS, RECYCLE TESTS WITH THIRD GENERATION (VENTURI) CONTACTOR, WITH STEAM ADDITION

Test	26	27	28	LT–01A	LT–01B	LT–01C
Additive	None	None	HCl	None	None	None
Sorbent Data						
Fresh Ca/S Mole Ratio (a)	1.41	1.26	1.27	1.28	1.28	1.26
Fresh Feedrate, lb/hr (b)	7.33	6.56	6.62	6.65	6.63	6.58
Recycle Feedrate, lb/hr	53.16	53.55	54.08	54.31	54.20	53.74
Recycle Ratio, lb recycle/lb fresh lime	7.25	8.16	8.17	8.17	8.17	8.17
Recycle Ratio, dry basis	6.74	6.69	6.69	6.69	6.70	6.69
Recycle Available Ca/S, Mol Ratio (a)	–	–	–	–	–	–
Total Available Ca/S, Mol Ratio (a)	–	–	–	–	–	–
Water Addition, lb/hr	0.00	5.96	6.02	6.05	6.04	5.99
lb Water/lb Recycle Sorbent	0.00	0.12	0.12	0.12	0.12	0.12
Duct Flue Gas Conditions						
In–Duct Residence Time, s	2.7	2.7	2.7	2.7	2.7	2.7
Duct Inlet SO_2 Content, ppmv–dry	1492	1497	1497	1497	1495	1500
Steam Addition, lb/hr	21	20	20	29	30	30
Approach to Saturation, °F						
Contactor Exit	–	<1	1	<1	0	0
Duct Exit	10	5	3	4	3	4
Baghouse Exit	13	14	19	6	4	4
Solids Loading, gr/scf	20.8	20.6	20.8	20.9	20.9	20.7
Contactor Inlet Temp, °F	288	287	286	282	283	281
Contactor Exit Temp, °F	141	139	137	138	138	140
Duct Exit Temp, °F	150	143	142	142	142	142
Baghouse Exit Temp, °F	152	151	157	143	142	141
Baghouse Exit Wet Bulb, °F	139	137	138	137	138	137
Duct Inlet Flue Gas Flow, scfm	340	340	340	340	340	340
SO_2 Removal, %						
In–Duct	77	80	86	82	72	67
System (Duct + Baghouse)	90	91	91	97	94	92
Sorbent Utilization, % (c)	64	72	72	76	73	73

(a) Includes all calcium in fresh and recycle sorbent not tied up with sulfur (e.g., $Ca(OH)_2$ and $CaCO_3$).
(b) Fresh feed was Mississippi hydrated lime.
(c) Based on flue gas analysis and fresh sorbent feed rate/composition, assuming steady state.

TABLE 23

SUMMARY OF RECYCLE TEST RESULTS WITH THIRD GENERATION (VENTURI) CONTACTOR, WITH STEAM ADDITION

Test	Fresh Ca/S, mol	Recycle Ratio, Dry	lb Water/ lb Recycle	Steam Add'n lb/hr	Approach, °F Contactor Exit	Approach, °F Duct Exit	Approach, °F Baghouse Exit	SO_2 Removal, % Duct	SO_2 Removal, % System	Sorbent Util.,(a) %
26	1.41	6.7	0	21	-	10	13	77	90	64
27	1.26	6.7	0.12	20	<1	5	14	80	91	72
28	1.27	6.7	0.12 (b)	20	1	3	19	86	91	72
LT-01A	1.28	6.7	0.12	29	<1	4	6	82	97	76
LT-01B	1.28	6.7	0.12	30	0	3	4	72	94	73
LT-01C	1.26	6.7	0.12	30	0	4	4	67	92	73

Common Conditions: Venturi throat nozzles - 5 gpm
Venturi pressure drop - 8" water

(a) Based on flue gas analysis and fresh sorbent feed rate/composition, assuming steady state.

(b) The recycle treatment water contained a small concentration (0.004 wt %) of hydrochloric acid (HCl).

TABLE 24

ANALYSES OF GRAB SAMPLES, PRODUCT FROM PUG MILL TEST

Drum (a):	1	1	2	2	3	3
Moisture, wt %	9.6	9.4	8.3	8.7	8.9	8.8
CaO, wt %	44.3	45.3	46.0	45.7	46.4	45.9
Sulfur, wt %	13.4	13.4	14.2	13.8	14.0	14.3
CO_3, wt %	6.4	6.4	6.6	6.3	6.7	6.8

(a) A sample was taken from the top and bottom of each of the three drums.

TABLE 25

RECYCLE TEST RESULTS: COMPARISON OF PUGMILL WITH HIGH INTENSITY MIXER FOR TREATING RECYCLE SORBENT

Test	Mixer	Fresh Ca/S, mol	Recycle Ratio (a)	lb Water/ lb Recycle Sorbent	Approach, °F		SO_2 Removal, %		Sorbent Util., % (b)
					Duct	Baghouse	Duct	System	
PT-1	Pug Mill	1.4	5.5	0.12	4	10	86	99	68
PT-2	High Intensity	1.4	5.5	0.12	3	9	85	98	67

(a) lb dry recycle/lb fresh lime.

(b) Based on baghouse ash analysis.

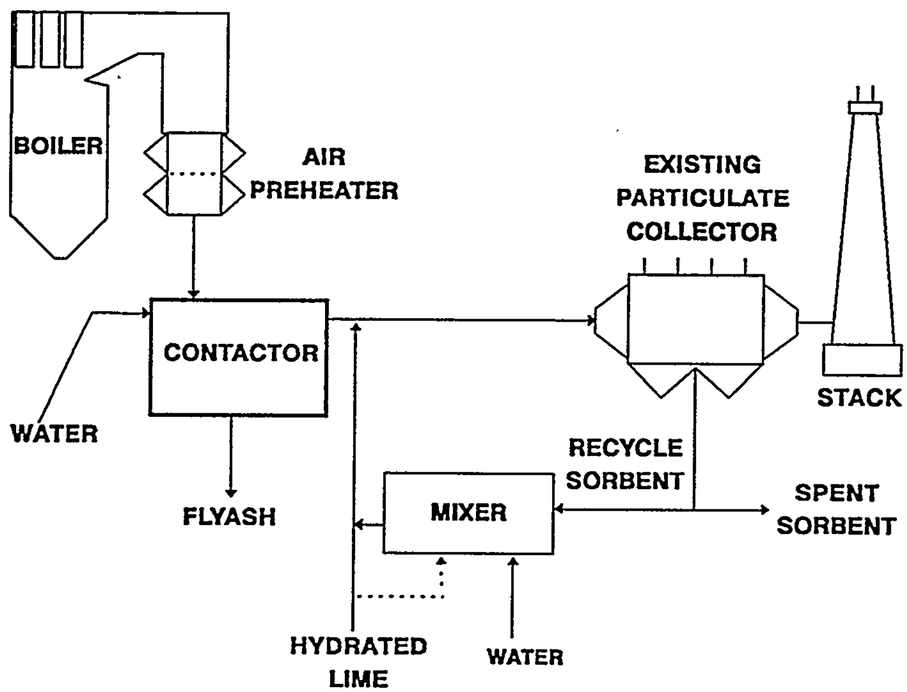

Figure 1. Schematic of Advanced Coolside Process.

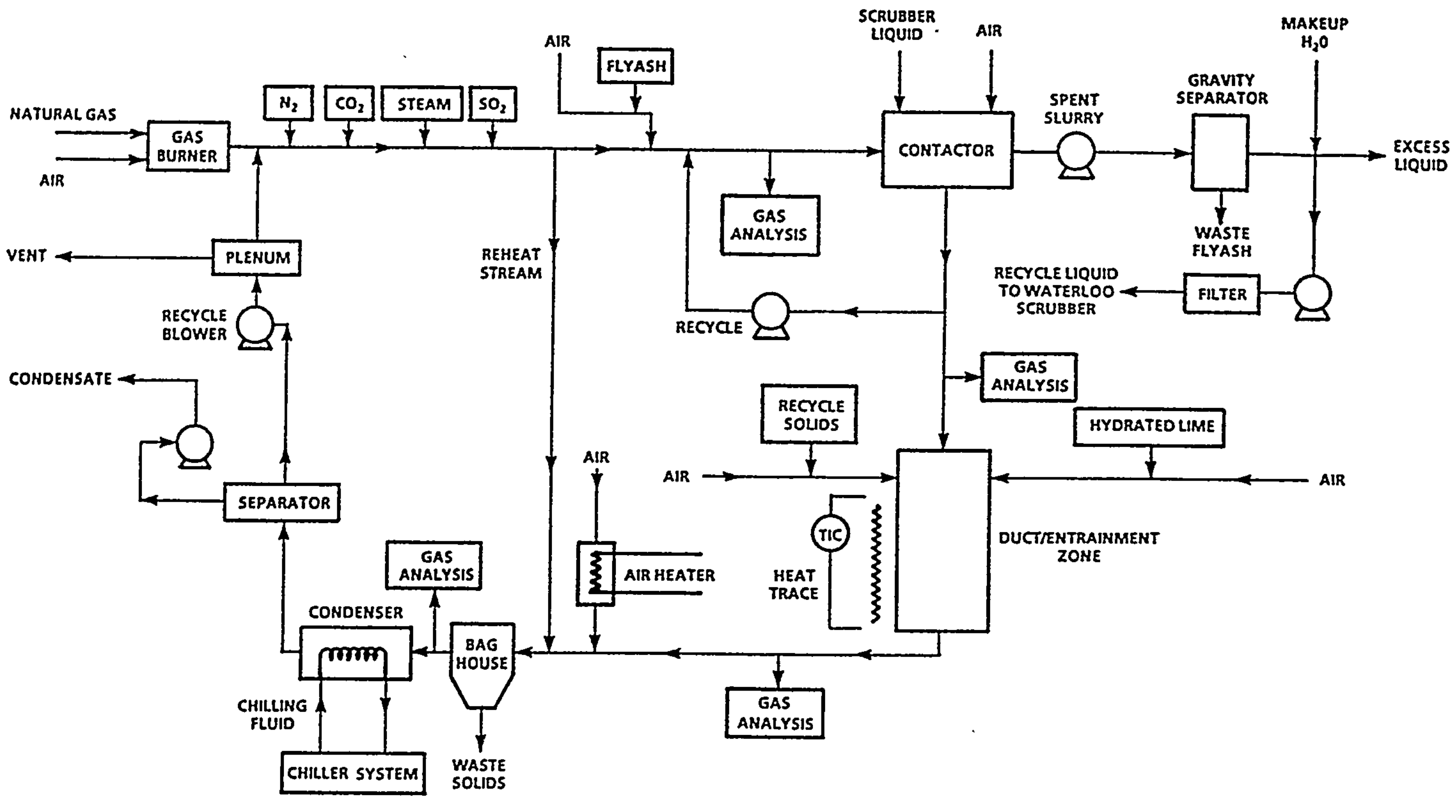

Figure 2. Schematic of Advanced Coolside Pilot Plant.

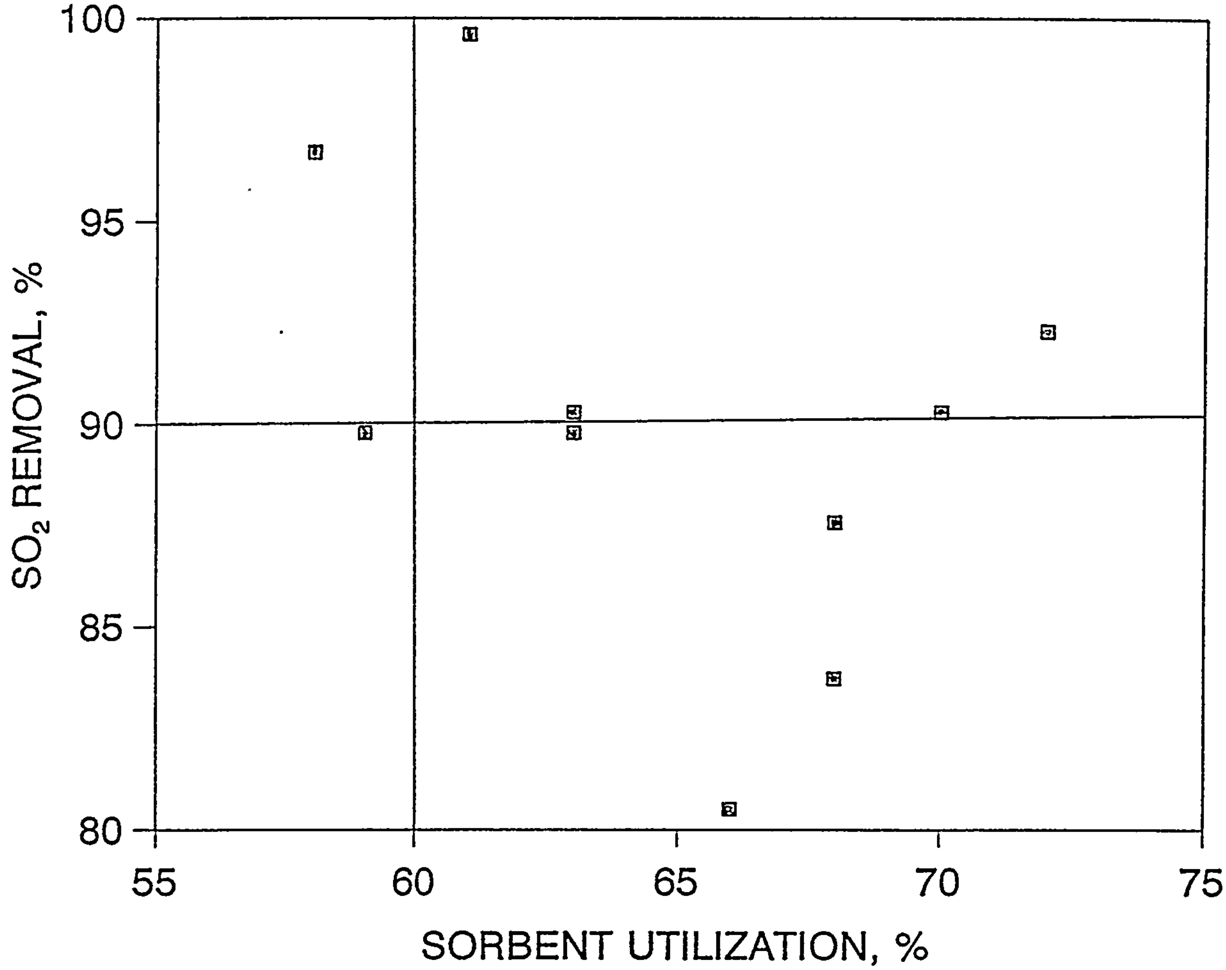

Figure 3. Recycle Simulation Tests: SO_2 Removals and Corresponding Sorbent Utilizations for Tests with Moisture Addition to the Recycle Sorbent.

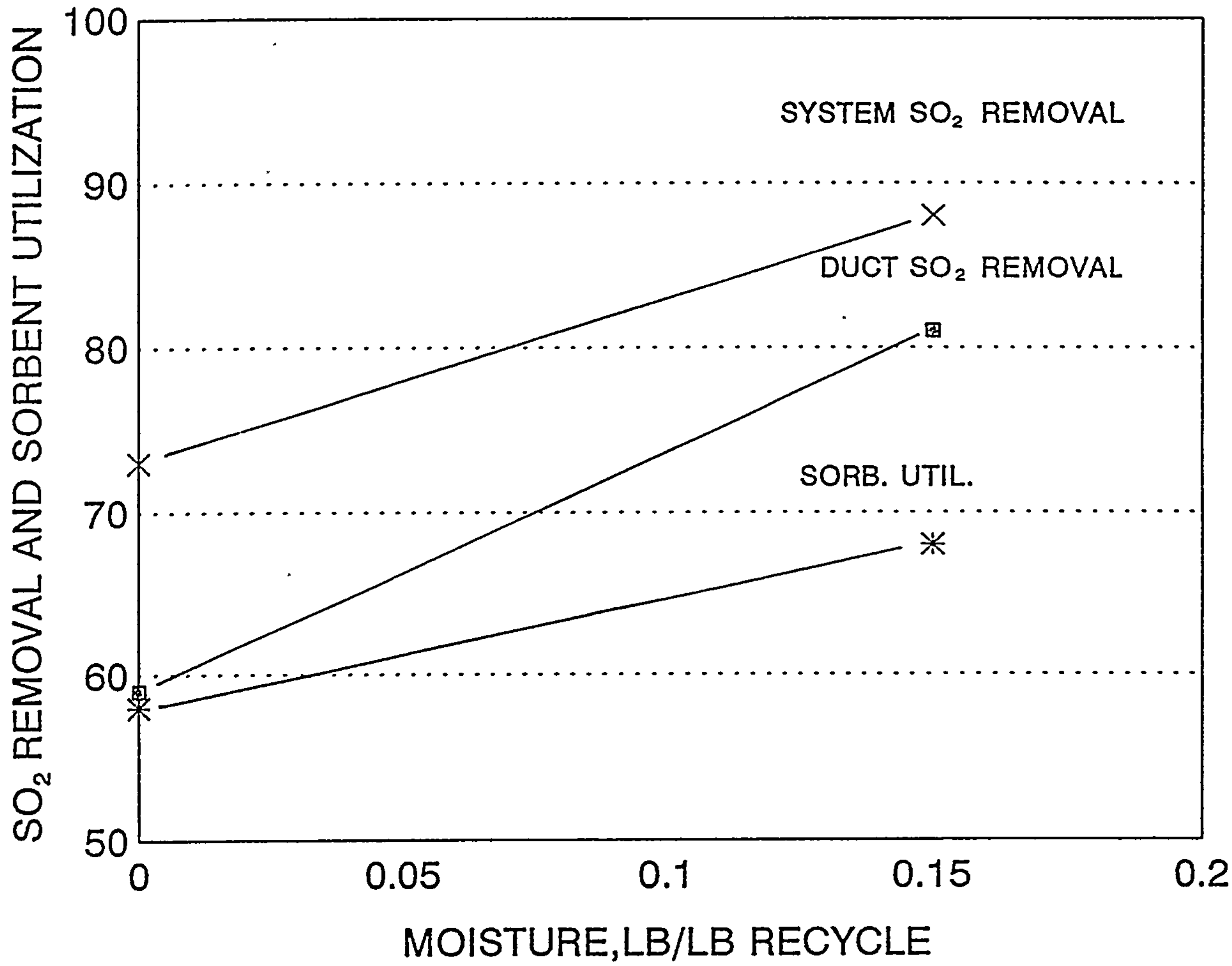

Figure 4. SO_2 Removals and Sorbent Utilizations, at a 1.2 Fresh Ca/S mol Ratio, a 5.0 Recycle Ratio and a 10 °F Baghouse Approach, as a Function of Moisture in the Recycle Sorbent

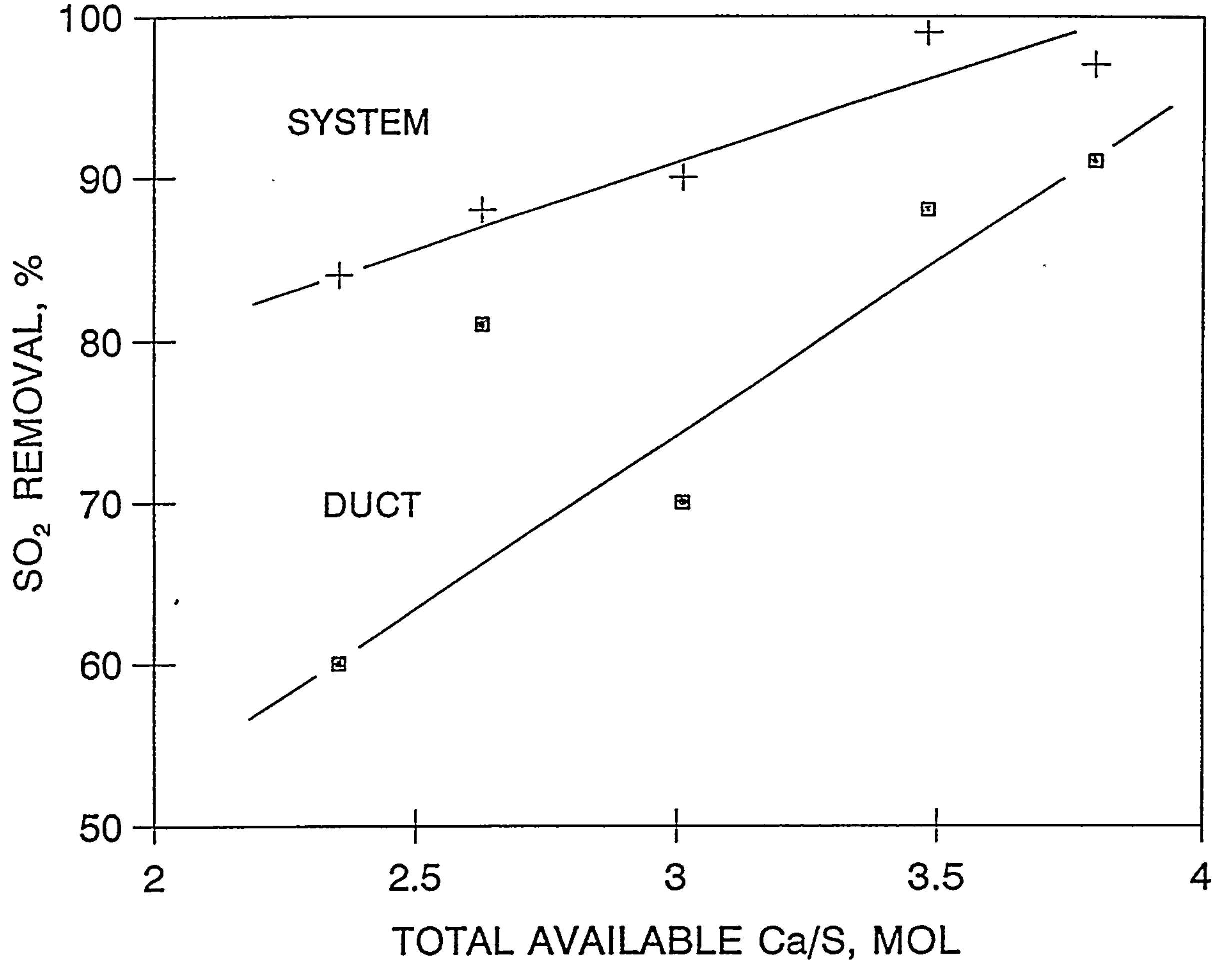

Figure 5. Duct and System SO_2 Removals, at a 10 °F Baghouse Approach and with 0.15 lb Water/lb Recycle Sorbent, as a Function of the Total Available Ca/S mol Ratio.

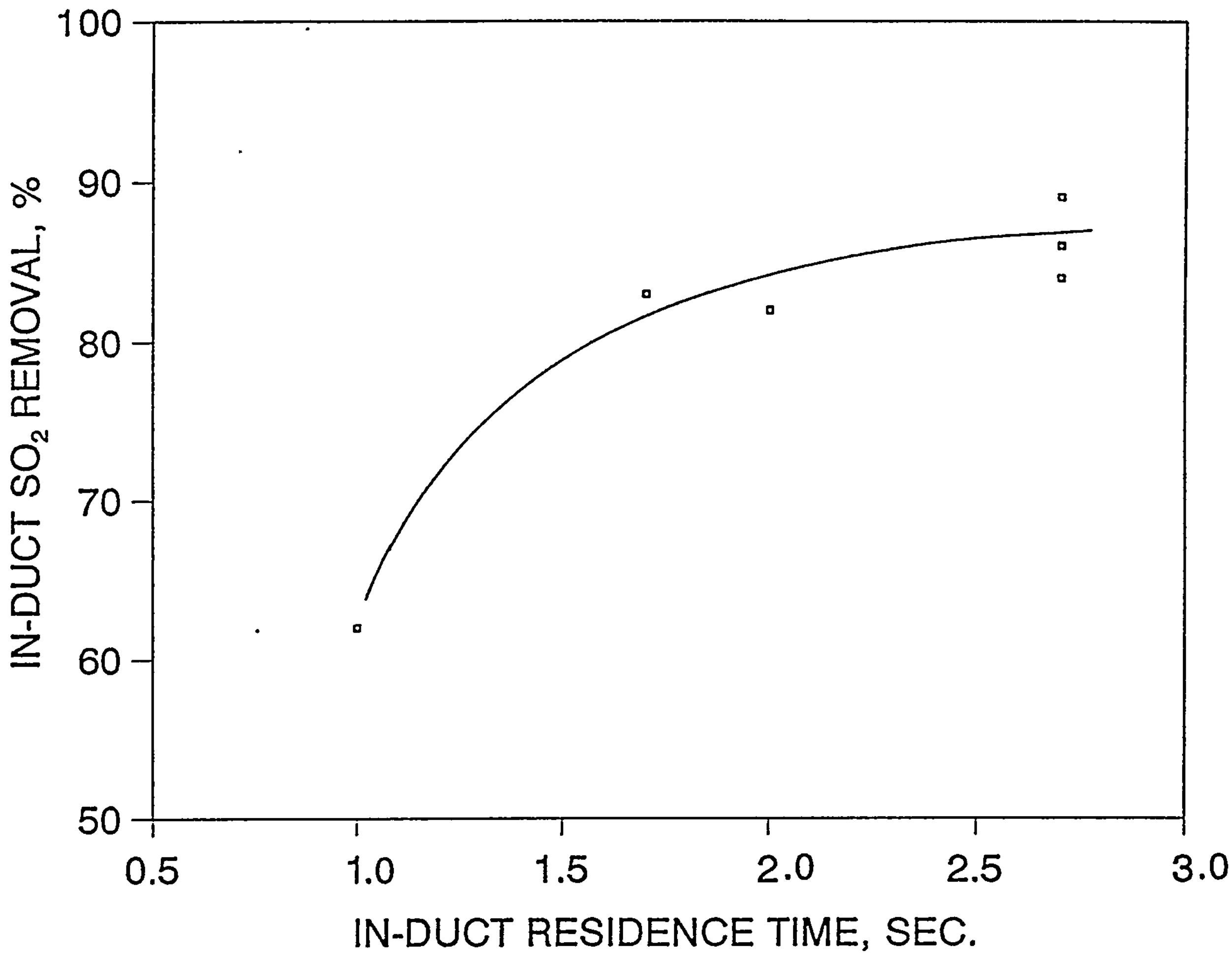

Figure 6. Effect of Residence Time on SO_2 Removal in Advanced Coolside Pilot Plant. (Recycle Simulation Test 13.)

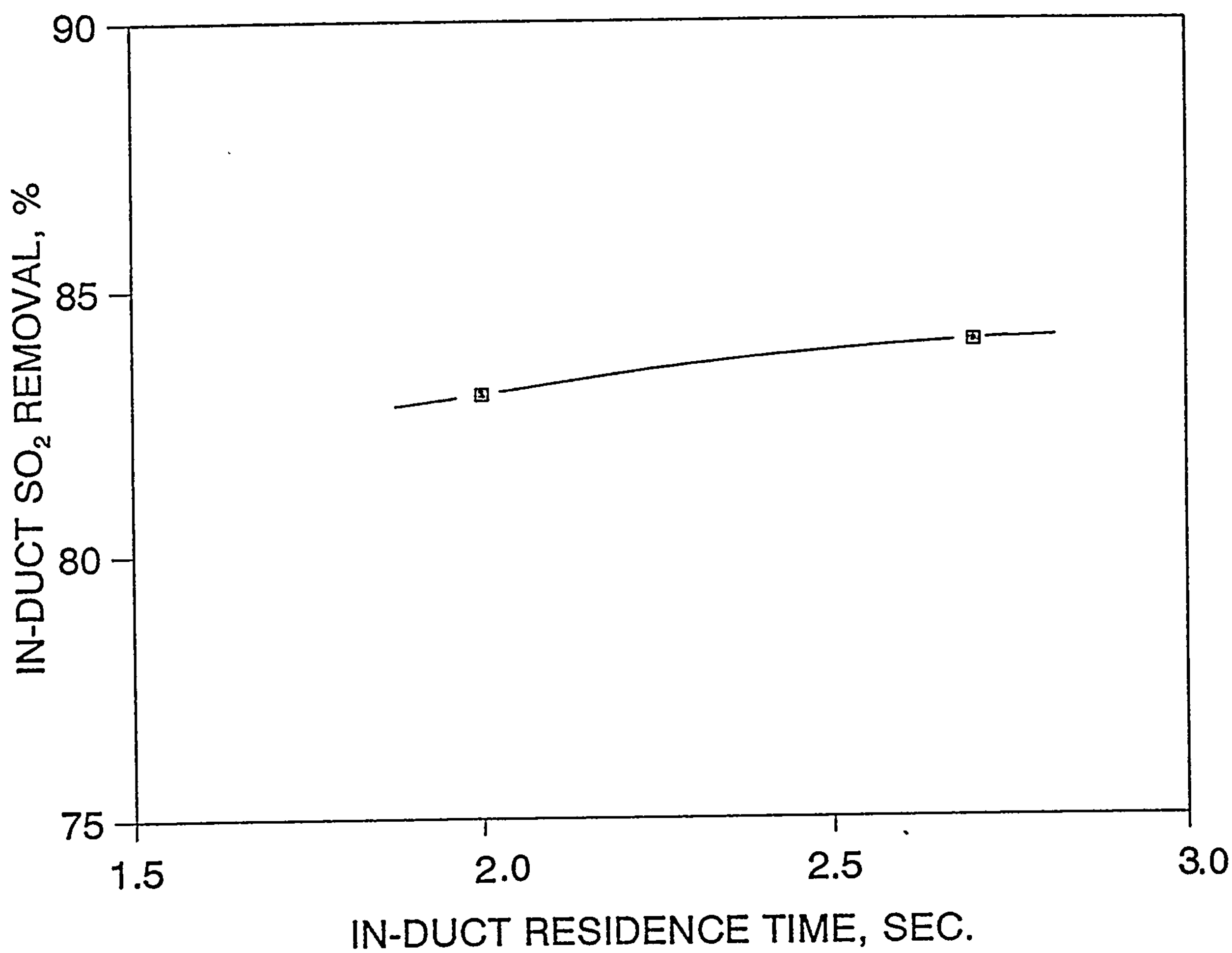

Figure 7. Effect of Residence Time on SO_2 Removal in Advanced Coolside Pilot Plant. (Recycle Simulation Test 12A.)

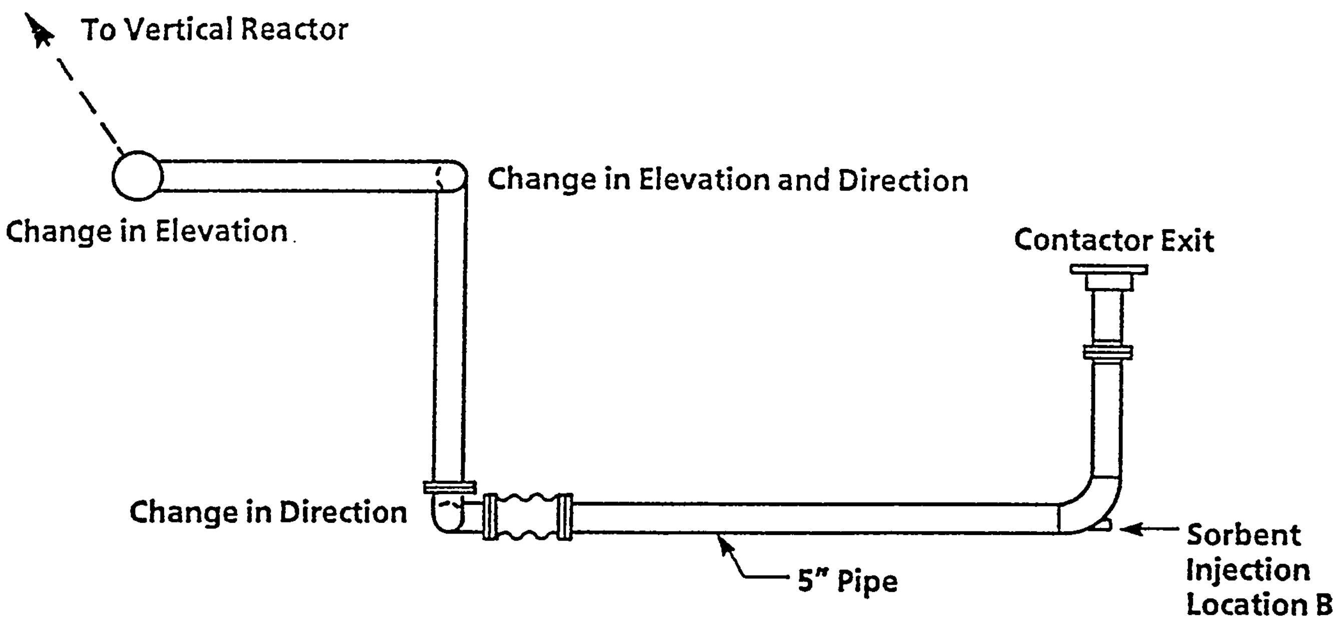

Figure 8. Configuration of Advanced Coolside Pilot Plant Ductwork.

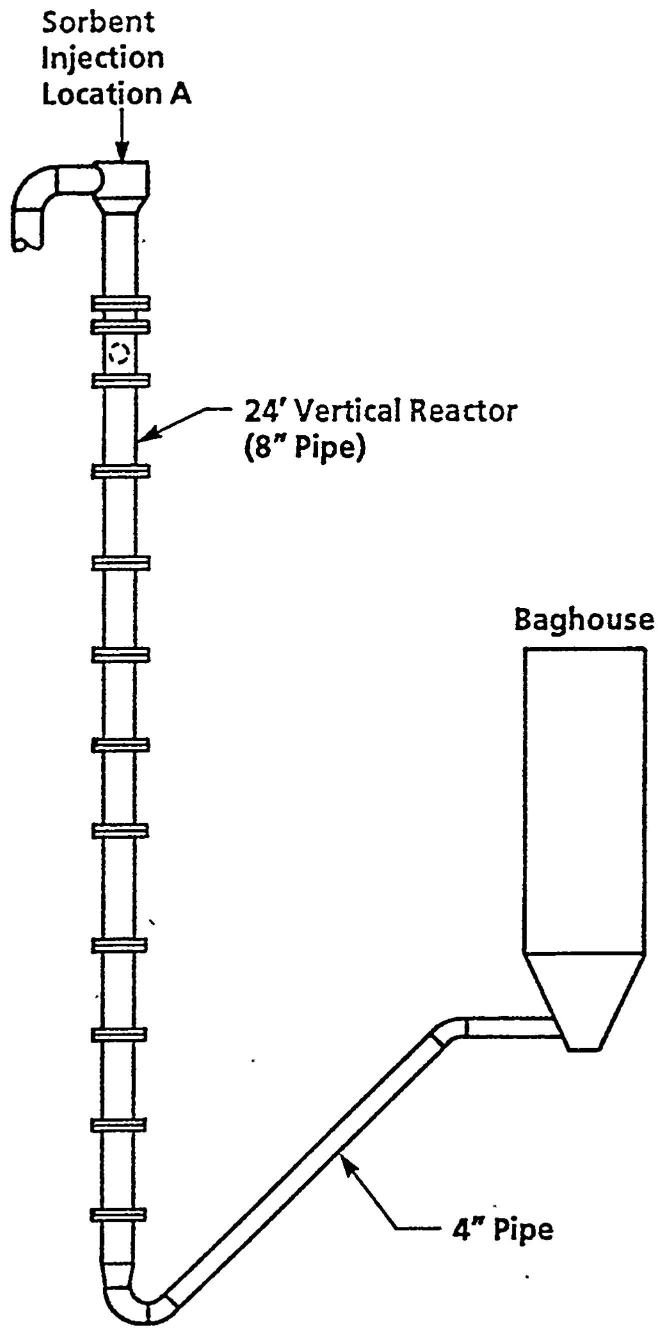

Figure 9. Configuration of Advanced Coolside Pilot Plant Ductwork.

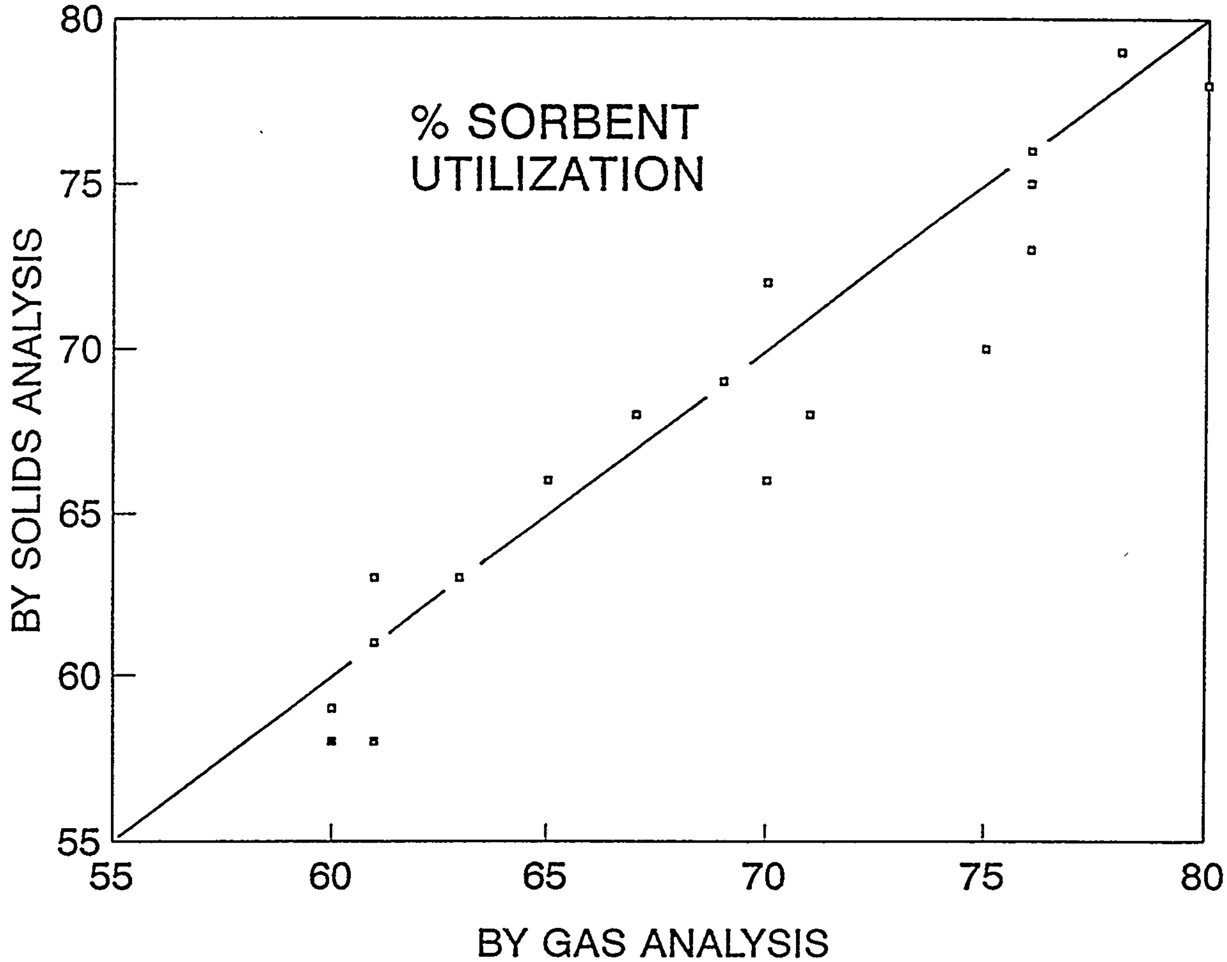

Figure 10. Comparison of Sorbent Utilizations Based on Solids Analyses with Utilizations Based on Gas Analyzers.

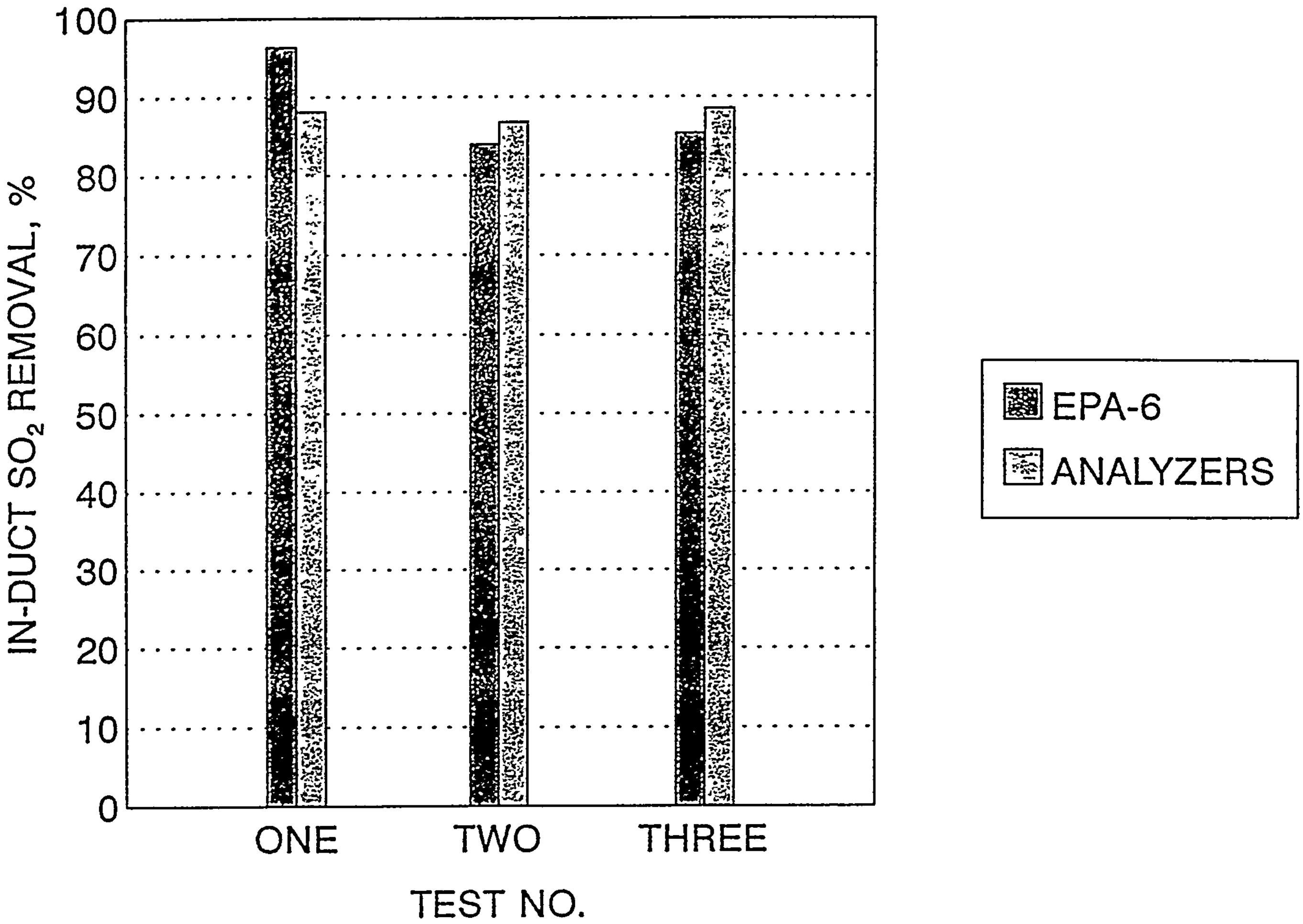

Figure 11. Comparison of EPA Method 6 and Flue Gas Analyzers: In-Duct SO_2 Removal in Test 11A.

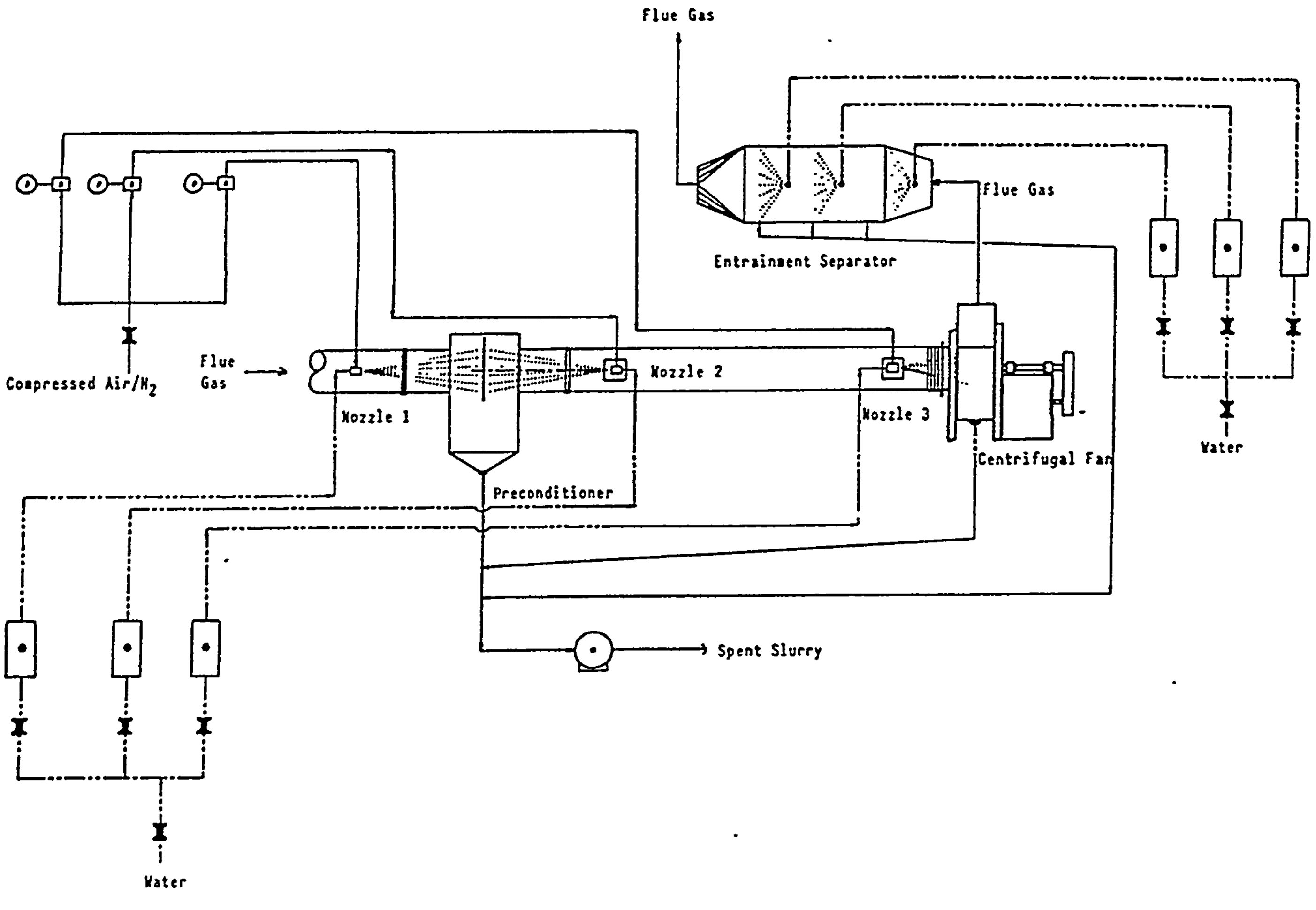

Figure 12. Schematic of First Generation Contactor.
(Waterloo Scrubber, Supplied by Turbotak, Inc.)

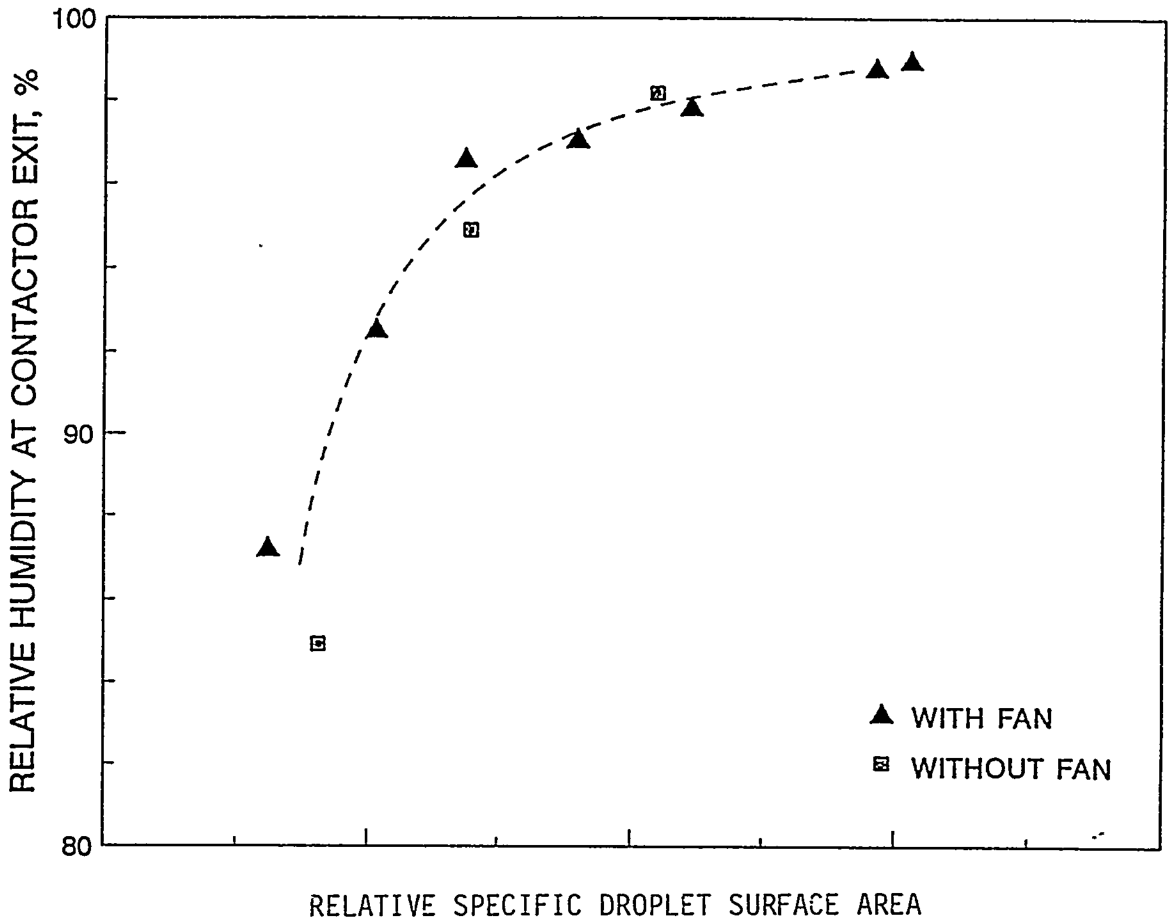

Figure 13. Humidification Tests Showing Similar Results With and Without the Scrubber Fan.

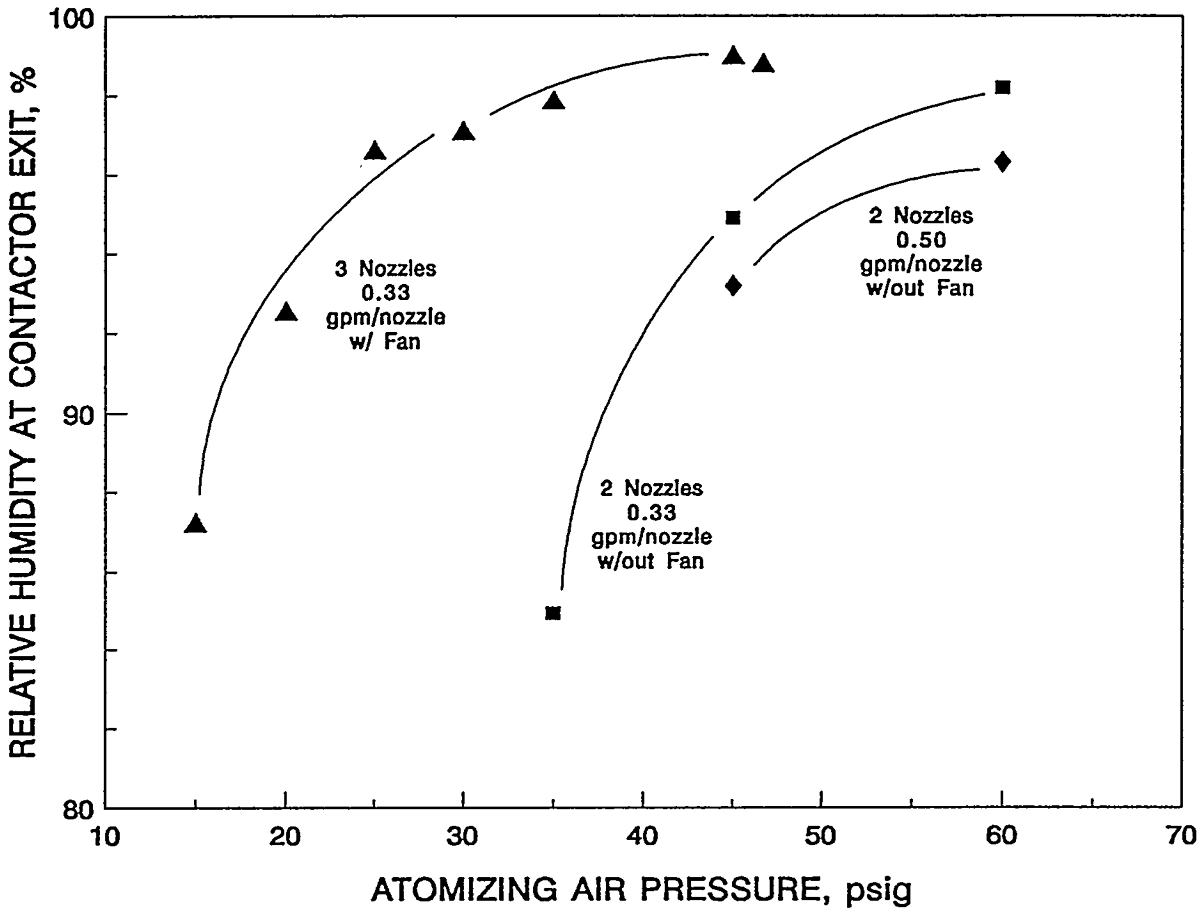

Figure 14. Effect of Atomizing Air Pressure on Relative Humidity, Original Contactor.

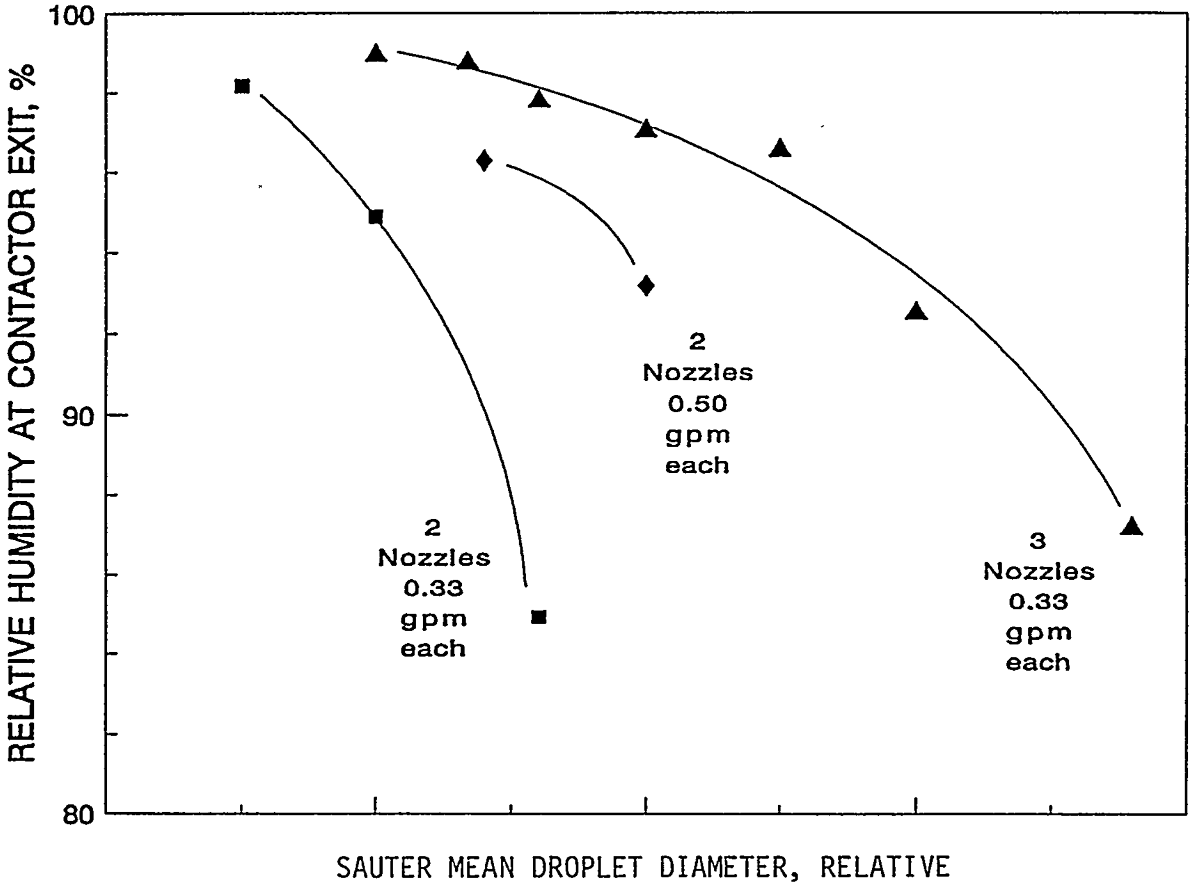

Figure 15. Effect of Sauter Mean Droplet Diameter on Relative Humidity, Original Contactor.

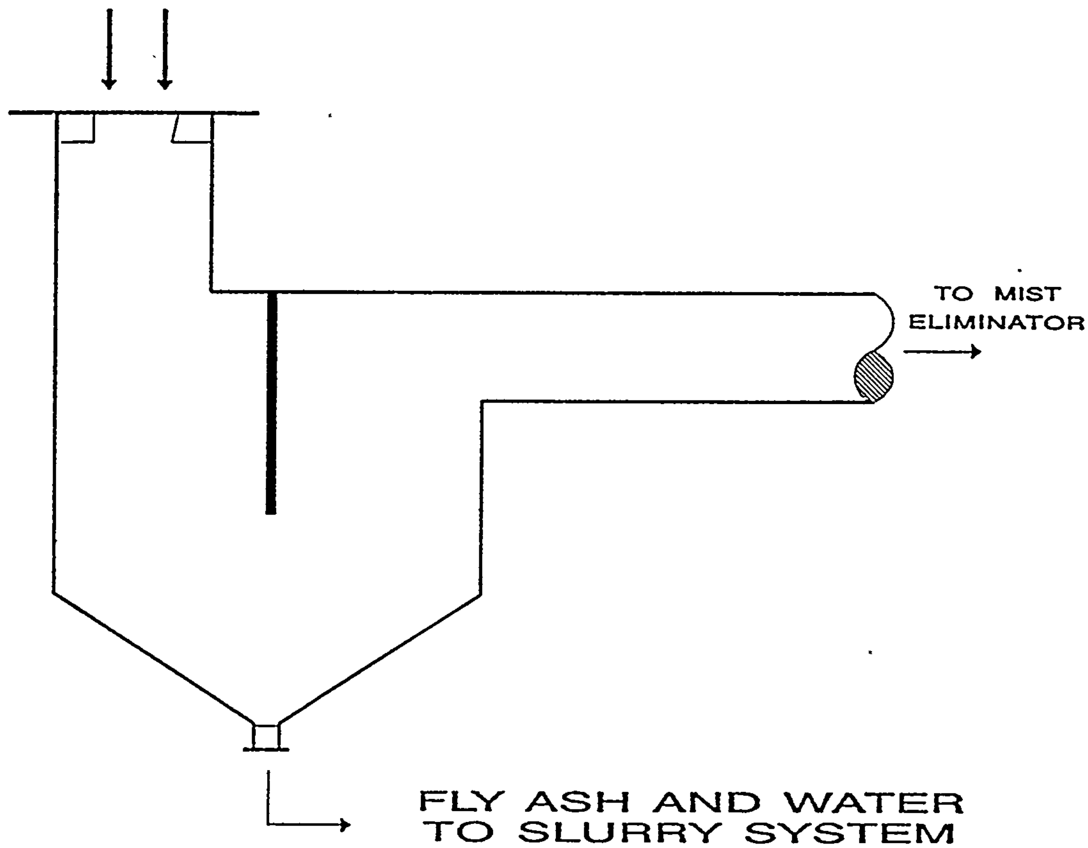

Figure 16. Schematic of Second Generation Contactor.

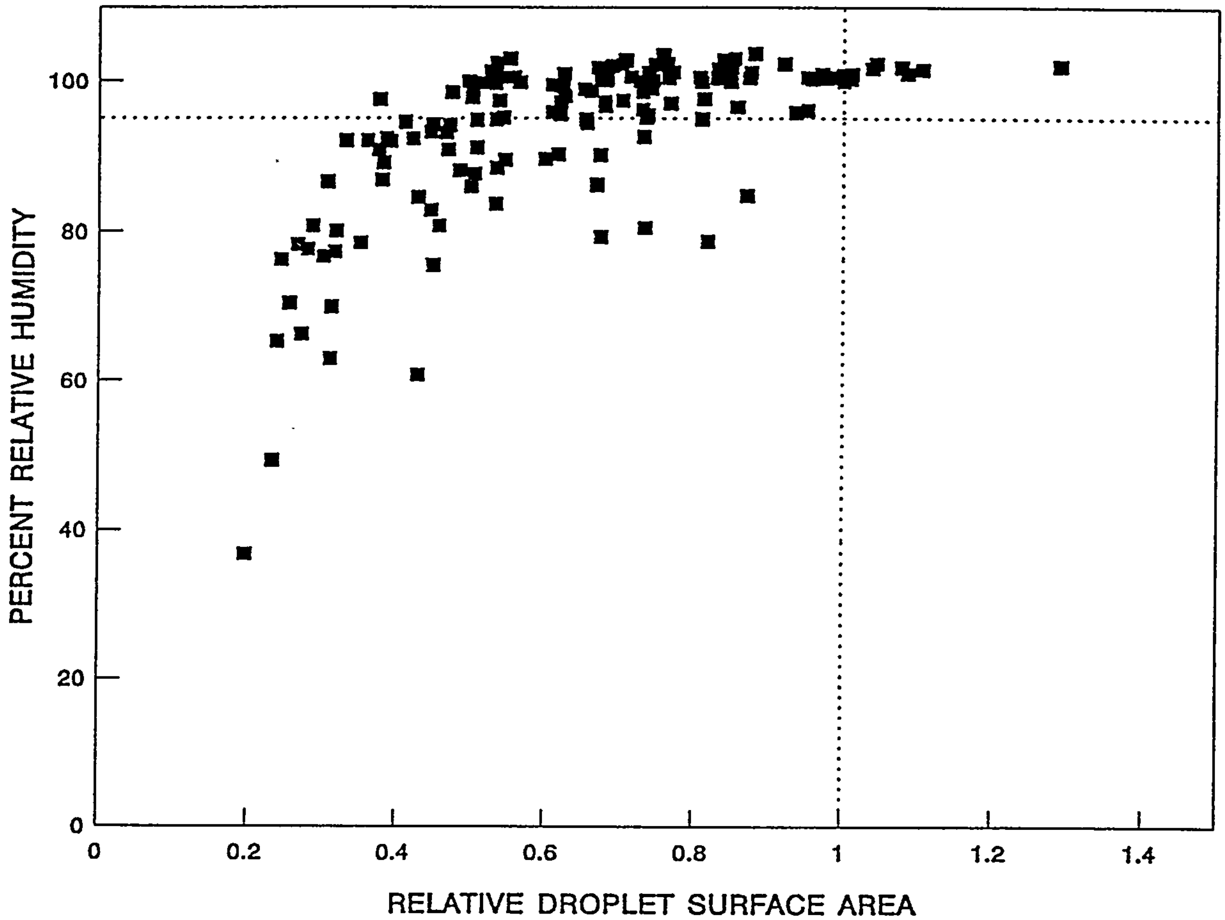

Figure 17. Saturation Efficiency for 150 Tests Using the Simplified Contactor.

Relative droplet surface area is the test droplet surface area (m^2/m^3 flue gas) divided by the droplet surface area produced when the nozzles are operated at the design air pressures and water flow rates.

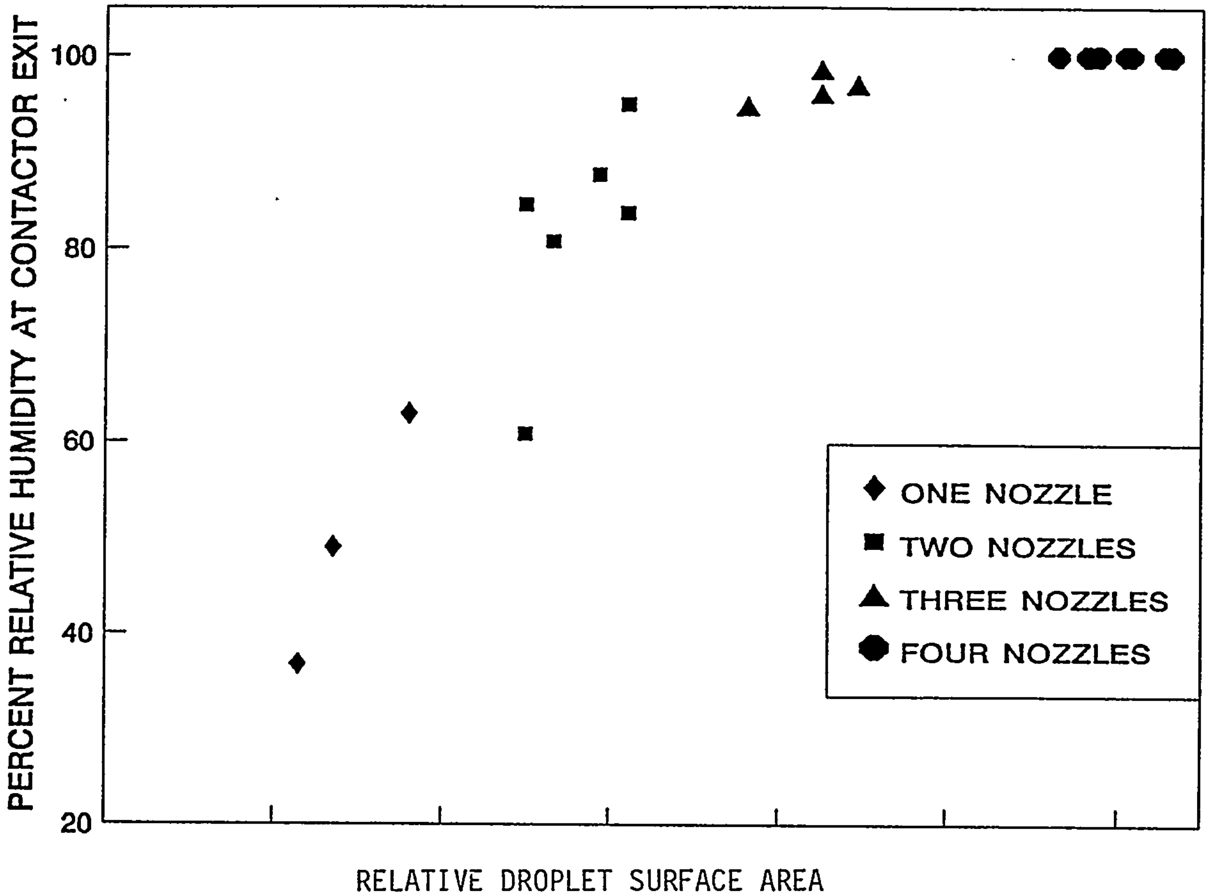

Figure 18. Humidification Tests Using 1, 2, or 3 Spray Nozzles.

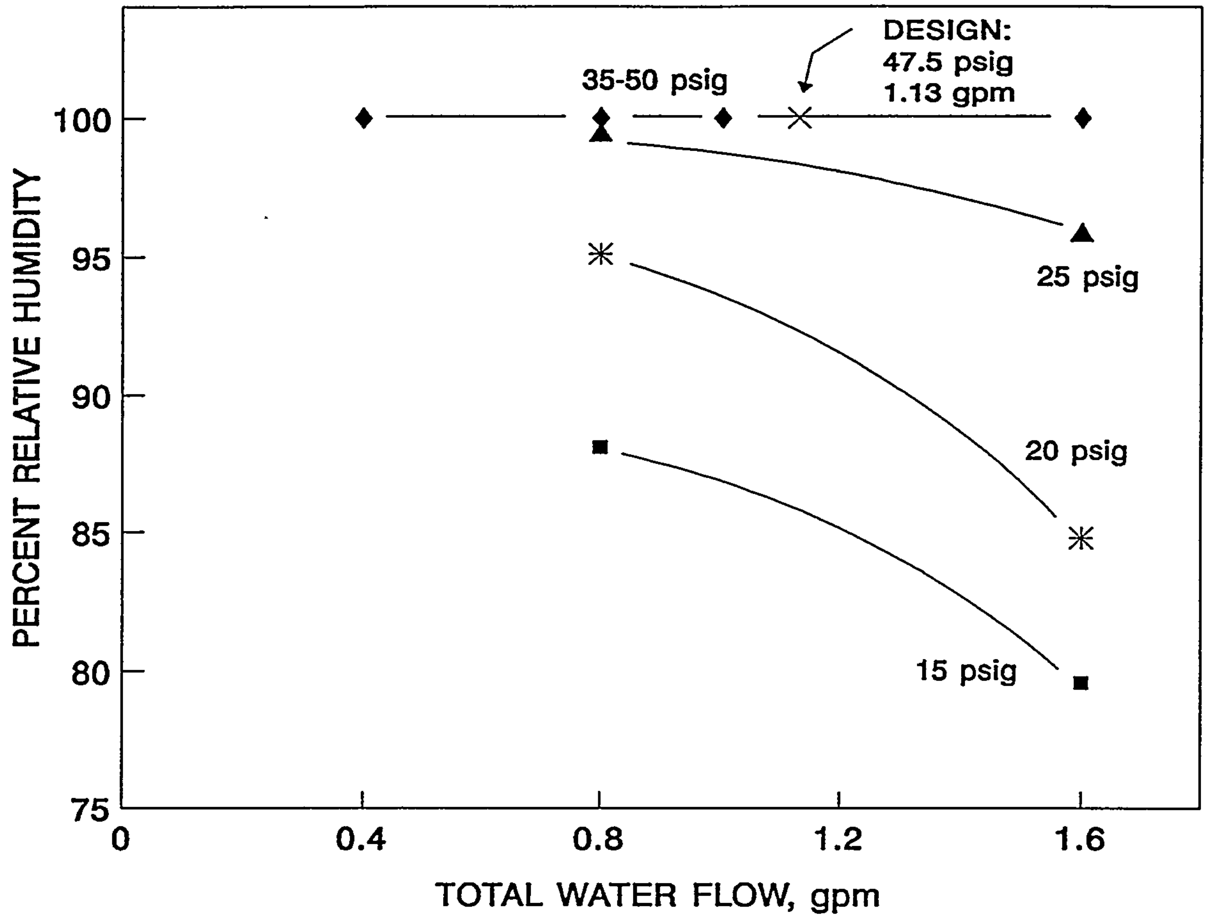

Figure 19. Effect of Atomizing Air Pressure and Water Flow on Humidification Using All Four Spray Nozzles, 500-525 scfm Gas Flow, Second-Generation Contactor.

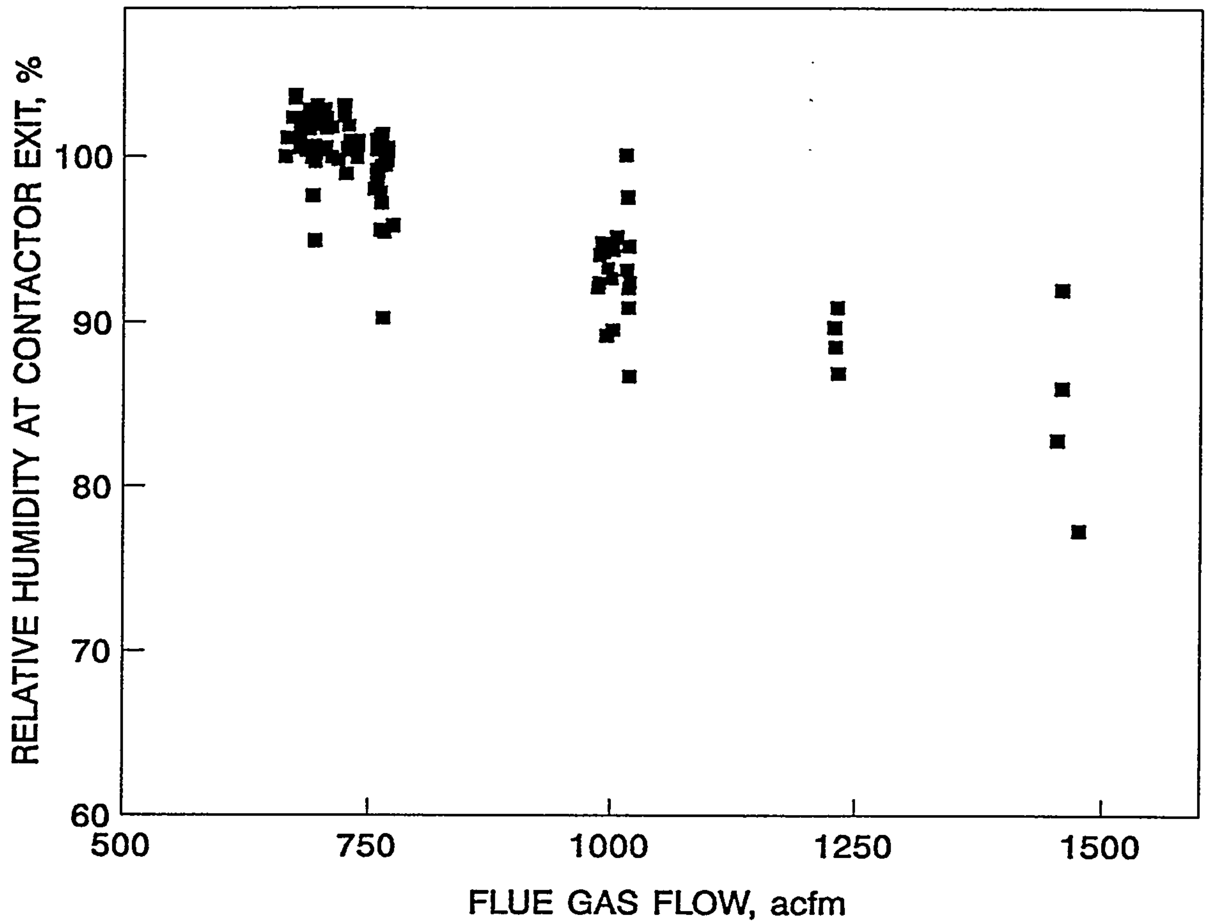

Figure 20. Effect of Contactor Flue Gas Throughput on Humidification Efficiency, Second Generation Contactor.

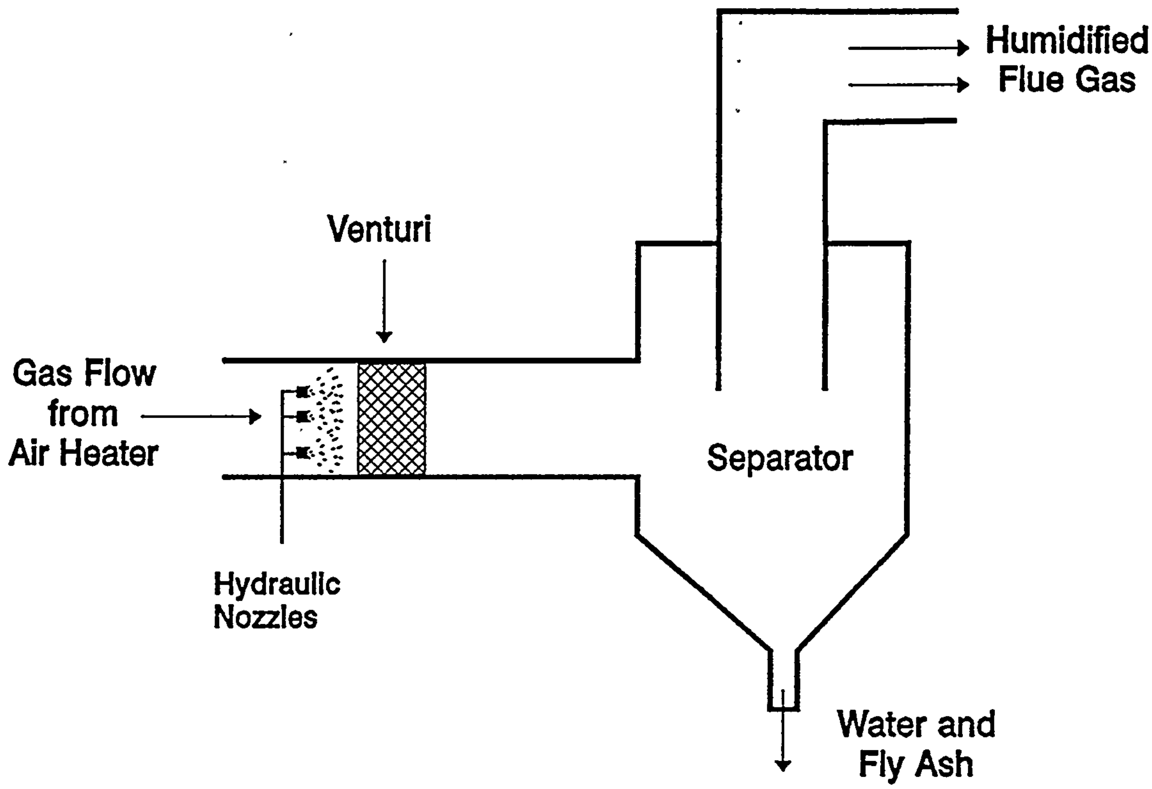

Figure 21. Schematic of Third Generation Contactor.

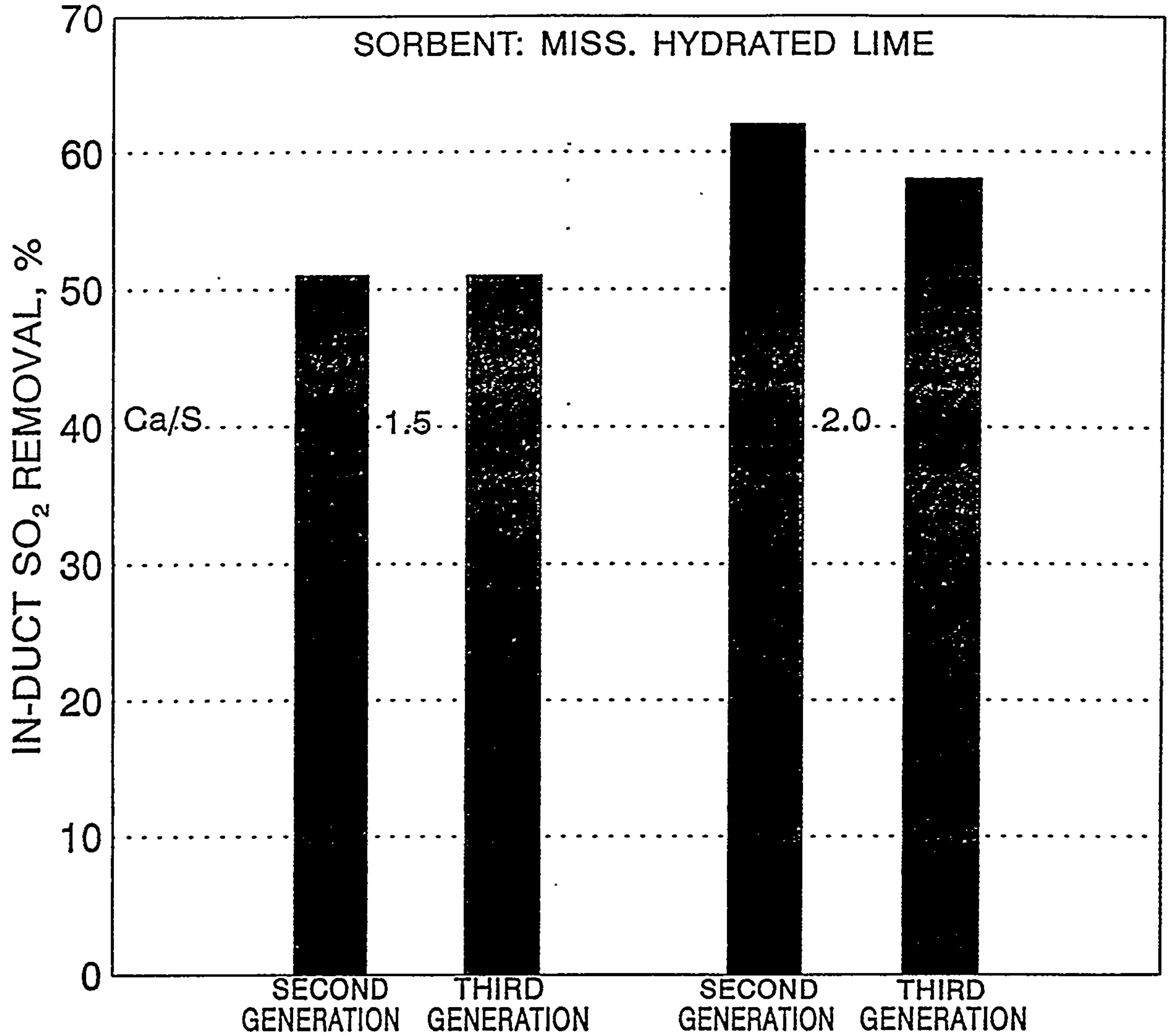

Figure 22. Comparison of Second Generation Contactor (Spray Chamber + Mist Eliminator) and Third Generation Contactor (Venturi + Cyclonic Separator), Once-Through Pilot Plant Tests.